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MUNICIPAL WASTE COMBUSTORS - BACKGROUND INFORMATION FOR PROPOSED STANDARDS: COST PROCEDURES

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Cost Procedures for the costing of new and existing municipal waste combustor facilities and associated equipment are presented. Cost procedures are developed for combustors, heat recovery equipment, humidification equipment, air pollution control devices for the reduction of particulate matter and acid gas emissions, and continuous emission monitoring equipment.

Costs in this report are divided into capital costs, operating and maintenance costs, and annualized costs. Costs associated with retrofitting existing facilities are also presented.

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FINAL REPORT

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1.0 INTRODUCTION

This report documents the development of cost procedures for costing new and existing municipal waste combustor (MWC) facilities, associated heat recovery equipment, humidification equipment, air pollution control devices (APCD's) for the reduction of particulate matter (PM) and acid gas emissions, and continuous emission monitoring equipment. Costs presented in this report are divided into three major cost categories:

- Capital Costs;
- Operating and Maintenance (O&M) Costs; and
- Annualized Costs (total O&M cost plus capital-related annual charges).

Each of these cost categories is further subdivided into individual cost elements.

Capital cost elements include direct costs (purchase equipment and installation costs), indirect costs, and contingencies. Direct costs consist of the basic and auxiliary equipment, the labor and material required to install the equipment, plus site preparation and buildings costs. Indirect costs are those costs which are not attributable to specific equipment items such as engineering, construction and field expenses, contractor fees, and start-up and performance tests. Contingencies cover any unpredicted events and other unforeseen expenses that may arise.

The O&M cost elements include direct and indirect costs. Direct O&M costs consist of operating and maintenance labor, fuel, utilities, materials and spare parts, supplies, waste disposal, and chemicals. These costs are dependent on the combustor capacity utilization. Indirect O&M costs, on the other hand, are totally independent of capacity utilization. These costs include plant and payroll overhead, real estate and local taxes, insurance, administrative charges, and replacement parts.

Total annualized costs are the sum of the direct and indirect O&M costs and capital recovery costs. Capital recovery costs are determined by multiplying the total capital costs by the capital recovery factor, which is based on the assumed interest rate and economic equipment life. A 10 percent real interest rate and a 15-year equipment life are assumed for the combustors and control equipment in this report. This translates into a capital recovery factor of 13.15 percent.

All costs are presented in December 1987 constant dollars. Chapter 2.0 of this report presents the costing procedures for new MWC plants. Included are procedures to estimate capital, operating and maintenance, and annualized costs for combustors, combustor-related equipment, flue gas temperature control, PM control using dry electrostatic precipitators (ESP's), acid gas control using either dry sorbent injection or spray dryers followed by a fabric filter for PM control, and continuous emission monitors (CEM's). Chapter 3.0 presents the costing procedures for existing MWC plants which include procedures used to estimate costs for operating existing combustors and for retrofitting emission controls. Emission controls evaluated include combustor modification, temperature control, PM control (rebuilding an existing ESP, adding plate area, or installing a new ESP), and acid gas controls using either dry sorbent injection or spray dryers with an existing ESP or a fabric filter for PM control.

Appendix A compares the costs between spray dryer/fabric filter and spray dryer/electrostatic precipitator systems applied to new plants. The purpose of this comparison was to determine whether (1) the costs of these systems differ sufficiently to warrant separate costing procedures for each system and (2) a single procedure can be used. Appendix B presents the tables summarizing the cost procedures presented in this report.

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COST COMPARISON BETWEEN SPRAY DRYER/FABRIC FILTERS (SD/FF) AND SPRAY DRYER/ELECTROSTATIC PRECIPITATOR (SD/ESP) SYSTEMS

A.1 INTRODUCTION

This appendix compares SD/FF and SD/ESP costs for two model mass-burn waterwall plants (a 250-tpd plant and a 3,000-tpd plant) at a PM outlet concentration of 0.01 gr/dscf. Costs presented in the appendix for SD/FF systems are based on cost procedures discussed in Section 2.4. Cost procedures presented in this appendix were used for SD/ESP. Lime requirements are based on a stoichiometric ratio of 1.5:1 for both systems. The objective of this comparison was to determine whether (1) the costs of these systems differ sufficiently to warrant separate costing procedures for each system and (2) a single procedure can be used.

A.2 COST COMPARISON BETWEEN SD/ESP'S AND SD/FF

Costs for SD/ESP's and SD/FF systems are estimated for two model mass-burn plants.[1] Model plant 1 is a 250-tpd plant with two combustors, whereas model plant 3 is a 3,000-tpd plant with four combustors. These plants were selected to cover the size range of most MWC facilities. For both plants, the SD systems are assumed to achieve 90 percent HCl and 70 percent SO_2 removal and an outlet PM emissions of 0.01 gr/dscf at 12 percent CO_2. The following two sections discuss the approach taken in estimating costs for SD/ESP applied to these model plants and the results of the cost comparison. The costs for SD/FF systems are based on procedures presented in Section 2.4 at a stoichiometric ratio of 1.5:1.

A.2.1 Approach Used to Estimate SD/ESP Costs

Table A-1 presents purchased equipment cost data for SD/ESP's provided by five manufacturers. The vendor quotations were based on design specifications for model mass-burn and refuse-derived fuel (RDF) plants. Because the costs in Table A-1 contain significant scatter, the costs for vendors A and C were used to develop the capital cost procedure for SD/ESP's applied to mass-burn combustors. Both manufacturers are experienced in SD technology. Furthermore, the costs reported by both were consistent and generally were conservative compared to the other vendor's costs. Limited

TABLE A-2. CAPITAL AND ANNUALIZED COSTS PROCEDURES FOR MASS BURN MWC's[a,b]

Capital Costs (dollars per ton/day of MSW processed)

1. Mass burn MWC without electrical generation:

   Unit Capital Costs = 50,420 $(430/\text{Size})^{0.39}$

2. Mass burn MWC with electrical generation:

   Unit Capital Costs = 60,700 $(430/\text{Size})^{0.39}$

3. Total Capital Costs = Unit Capital Cost * TPD

Annualized Costs

1. Operating and Maintenance Costs excluding waste disposal:

   For mass burn refractory wall MWC,
   Costs = (15.7 - 0.00115 TPD) * Total Capital Costs/100

   For mass burn waterwall MWC,
   Costs = (12.5 - 0.00115 TPD) * Total Capital Costs/100

2. Capital Recovery[c]

   Costs = CRF * Total Capital Costs

3. Waste Disposal of Bottom Ash:

$$\text{Costs} = \frac{1}{24} * \frac{100 - \text{WR}}{100} * \text{TPD} * \text{HRS} * \text{WDC}$$

---

[a] Costs are estimated in December 1987 dollars.

[b] Size = combustor MSW feed rate, tons/day
TPD = plant MSW feed rate, tons/day
HRS = hours of operation
CRF = Capital recovery factor, 0.1315 based on 10 percent interest rate and 15-year economic life
WR = weight reduction MSW in the combustor percent
WDC = waste disposal cost rate, dollars per ton (typically $25/ton)

[c] Applies only to new plants. Capital recovery costs are not estimated for retrofit applications, since the capital costs are sunk.

cost data were available from vendor E at other outlet PM emission levels to substantiate the relative high equipment cost at 0.01 gr/dscf at 12 percent CO_2.

Table A-2 presents the capital cost procedures for SD/ESP applied to mass-burn facilities only. A cost equation was developed relating purchased equipment costs in Table A-1 at an outlet PM emission level of 0.01 gr/dscf at 12 percent CO_2 with flue gas flowrate on a logarithmic basis. The resultant equipment cost equation updated to December 1987 dollars using the Chemical Engineering Plant Index is given below:

$$\text{Equipment Costs, } 10^3 \ \$ = 5.896 \ Q^{0.535}$$

where:

Q = 125 percent of the actual flue gas flowrate, acfm.

Both installation and indirect costs are 60 percent of the equipment costs. Assuming that the indirect costs are 33 percent of the direct costs,[2] the direct cost equation for SD/ESP system shown in Table A-2 can be derived. Total direct cost equations for ductwork and the I.D. fan for SD/FF systems in Section 2.4 are used directly for SD/ESP systems. To be consistent with the SD/FF procedures in Section 2.4, costs for installation, indirect capital costs, and contingencies for SD/ESP are based on the same percentages used in the SD/FF procedures.

Operating costs for SD/ESP were estimated using Table A-3. These operating costs are based on lower operating labor requirements (3 man-hours/shift versus 4 man-hours/shift) and lower fan gas-side pressure drop requirements (5.5 inches versus 12.5 inches) than those for SD/FF. The gas-side pressure drop of 5.5 inches is based on a pressure drop of 5 inches across the SD and 0.5 inches across the ESP. Electricity costs are included for ESP energization. Additional costs are included for the SD/FF systems for bag replacement and compressed air. The same cost rates used to estimate SD/FF operating costs in Section 2.4 are used for estimating operating costs for SD/ESP systems in December 1987 dollars.

A.2.2 Cost Comparison Results

Tables A-4 and A-5 present costs for both SD/ESP and SD/FF systems applied to 250- and 3,000-tpd mass-burn plants, respectively. e capital

TABLE A-2. CAPITAL COST PROCEDURES FOR SD/ESP ON NEW MASS-BURN PLANTS

Total Direct Costs (December 1987 dollars)[a]

Single SD/ESP Unit: Costs, $10^3\$ = 7.087\ (Q)^{0.535}$

Ductwork: Costs, $10^3\$ = 1.387 * L * Q^{0.5}/1000$

Fan: Costs, $10^3\$ = 1.875 * Q^{0.96}/1000$

Multiple Units: Multiply the above costs by the number of units.

Indirect Costs = 33% of total direct costs.

Contingency = 20% of sum of direct and indirect costs.

Total Capital Investment = Total Direct Costs + Indirect Costs + Contingency Costs.

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[a] Q = 125 percent of the actual flue gas flowrate, acfm

L = Duct length, feet

Cost procedures assume that the total installed costs are 133 percent of the total direct capital costs.[2]

TABLE A-3. ANNUAL OPERATING COSTS PROCEDURES FOR SD/ESP ON NEW MASS-BURN PLANTS

	Reference
Operating Labor: 3 man-hours/shift; $12/man-hour	3, 4
Supervision: 15% of operation labor costs	4
Maintenance:	
Labor -- 2 man-hour/shift 10% wage rate premium over operating labor wage	3, 4
Materials -- 2% of direct capital costs	3
Electricity: Electricity costs = $0.046/kwh	
ESP Energization -- 1.5 watts/ft^2 plate area	5
Fan -- 5.5 inches of water pressure drop	6, 7
Atomizer Auxiliary Equipment -- Kw = 6kw per 1,000 lbs/hr of slurry feed + 15kw	8
Pump -- 20 feet of pumping height 10 psi discharge pressure 10 ft/sec velocity in pipe	9
Water: Based on water flowrate required for cooling flue gas to 300°F and water cost rate of $0.50/1000 gal	10
Lime: Based on lime feed rate to the spray dryer calculated by assuming a stoichiometric ratio of 1.5:1. Apply appropriate lime costs in $/ton ($70/ton)	11
Solid Waste: Calculate solid waste disposal rate collected by the ESP and the spray dryer and apply appropriate tipping fee in $/ton. (Assume $25/ton)	12

TABLE A-3. ANNUAL OPERATING COSTS PROCEDURES FOR SD/ESP ON NEW MASS-BURN PLANTS (Continued)

		Reference
Overhead:	60% of the sum of all labor costs (operating, supervisory, and maintenance) plus materials	13
Taxes, Insurance, and Administrative Charges:	4% of total capital costs	13
Capital Recovery:	15-year life and 10% interest rate	14

TABLE A-4. COSTS FOR SD/ESP'S AND SD/FF'S FOR A 250-TPD MASS-BURN PLANT[a]

Model Plant No. 1

250 tpd Mass-Burn Facility with 2 Combustors
Outlet PM Emissions = 0.01 gr/dscf

	SD/FF	SD/ESP
Capital Cost (1,000 $)		
Total Direct	3,270	3,730
Total Indirect	1,080	1,230
Contingency	870	993
Total Capital Costs	5,220	5,960
Operating Costs (1,000$)		
Direct Costs:		
Operating Labor	96	72
Supervision	14	11
Maintenance Labor	53	40
Materials	65	75
Electricity	62	51
Water	1	1
Lime	50	50
Waste Disposal	81	81
Bag Replacements	15	0
Compressed Air	8	0
Indirect Costs:		
Overhead	137	119
Taxes, Insurance, & Administration	209	238
Total Operating Costs	791	738
Annualized Costs		
Capital Recovery	687	783
Total Annualized Costs	1,480	1,520

---

[a] Costs are in December 1987 dollars.

TABLE A-5. COSTS FOR SD/ESP'S AND SD/FF'S FOR A 3,000-TPD MASS-BURN PLANT[a]

Model Plant No. 3

3,000 tpd Mass-Burn Facility with 4 Combustors
Outlet PM emissions = 0.01 gr/dscf

	SD/FF	SD/ESP
Capital Cost (1,000 $)		
Total Direct	17,340	20,260
Total Indirect	5,720	6,690
Contingency	4,610	5,390
Total Capital Costs	27,600	32,300
Operating Costs (1,000 $)		
Direct Costs:		
Operating Labor	192	144
Supervision	29	22
Maintenance Labor	106	106
Materials	347	405
Electricity	629	504
Water	12	12
Lime	594	594
Waste Disposal	975	975
Bag Replacements	184	0
Compressed Air	98	0
Indirect Costs:		
Overhead	404	406
Taxes, Insurance, & Administration	1,110	1,290
Total Operating Costs	4,680	4,460
Annualized Costs		
Capital Recovery	3,640	4,250
Total Annualized Costs	8,320	8,710

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[a] Costs are in December 1987 dollars.

costs for SD/ESP systems are higher than those for SD/FF systems for both plants. This is because ESP capital costs are more sensitive to PM removal requirements than those for FF's. At the removal efficiencies required to achieve an outlet loading of 0.01 gr/dscf, the capital costs for a SD/ESP are roughly 15 percent higher than for a SD/FF.

Tables A-4 and A-5 show that operating costs for SD/ESP and SD/FF systems are essentially the same. For both plants, capital-related operating costs are greater for an SD/ESP than for an SD/FF. The noncapital-related costs for an SD/ESP are lower. The magnitude of these cost differences are roughly equal, resulting in about the same operating costs for both SD systems.

Because of lower capital costs, annualized costs for SD/FF systems are roughly 4 percent less than SD/ESP systems for both model plants. The results from this cost comparison, which showed the annualized costs for both systems are similar, agreed with those presented in another cost study prepared for the State of New York.[15]

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11. Chemical Marketing Reporter. Volume 233. Number 1. January 4, 1988.

12. Reference 6, p. 2-29.

13. Reference 6, p. 2-31.

14. Reference 2, pp. 2-17 and 2-18.

15. Reference 7, pp. 6-1 to 6-17.

TABLE B-1. CAPITAL AND ANNUALIZED COST PROCEDURES FOR MODULAR MWC's[a,b]

Capital Costs

1. Modular MWC without heat recovery:

   Unit Capital Cost = $24,300 per ton/day of MSW processed

2. Modular MWC producing steam (without generating electricity):

   Unit Capital Cost = $32,500 per ton/day of MSW processed

3. Modular MWC generating electricity:

   Unit Capital Cost = $54,600 per ton/day of MSW processed

4. Total Capital Costs = Unit Capital Costs * TPD

Annualized Costs[b]

1. Operating and Maintenance Costs excluding waste disposal:

   For TPD < 150 and HRS < 6,000,
   Costs = (10 - 0.23 TPD + 0.006 HRS) * Total Capital Costs/100

   Otherwise,
   Costs = (15.7 - 0.00115 TPD) * Total Capital Costs/100

2. Capital Recovery[c]:

   Costs = CRF * Total Capital Costs

3. Waste Disposal of Bottom Ash:

$$\text{Costs} = \frac{1}{24} * \left(\frac{100 - WR}{100}\right) * TPD * HRS * WDC$$

---

[a] Costs are estimated in December 1987 dollars.

[b] TPD = plant MSW feed rate, tons/day
HRS = hours of operation
CRF = capital recovery factor, 0.1315 based on 10 percent interest rate and 15-year economic life
WR = weight reduction of MSW in the combustor, percent
WDC = waste disposal cost rate, dollars per ton (typically $25/ton)

[c] Applies only to new plants. Capital recovery costs are not estimated for retrofit applications since the capital costs are sunk.

TABLE B-2. CAPITAL AND ANNUALIZED COSTS PROCEDURES FOR MASS-BURN MWC's[a,b]

Capital Costs (dollars per ton/day of MSW processed)

1. Mass-burn MWC without electrical generation:

   Unit Capital Costs = $50{,}420\ (430/\text{Size})^{0.39}$

2. Mass-burn MWC with electrical generation:

   Unit Capital Costs = $60{,}700\ (430/\text{Size})^{0.39}$

3. Total Capital Costs = Unit Capital Cost * TPD

Annualized Costs

1. Operating and Maintenance Costs excluding waste disposal:

   For mass-burn refractory wall MWC,
   Costs = (15.7 - 0.00115 TPD) * Total Capital Costs/100

   For mass-burn waterwall MWC,
   Costs = (12.5 - 0.00115 TPD) * Total Capital Costs/100

2. Capital Recovery[c]

   Costs = CRF * Total Capital Costs

3. Waste Disposal of Bottom Ash:

   $$\text{Costs} = \frac{1}{24} * \frac{100 - \text{WR}}{100} * \text{TPD} * \text{HRS} * \text{WDC}$$

---

[a] Costs are estimated in December 1987 dollars.

[b] Size = combustor MSW feed rate, tons/day
TPD = plant MSW feed rate, tons/day
HRS = hours of operation
CRF = Capital recovery factor, 0.1315 based on 10 percent interest rate and 15-year economic life
WR = weight reduction MSW in the combustor percent
WDC = waste disposal cost rate, dollars per ton (typically $25/ton)

[c] Applies only to new plants. Capital recovery costs are not estimated for retrofit applications, since the capital costs are sunk.

TABLE B-3. CAPITAL AND ANNUALIZED COST PROCEDURES FOR RDF FACILITIES[a,b]

Capital Costs (dollars per ton/day of RDF processed)

1. Coarse RD facility:

   Unit Capital Costs = 73,600 $(400/\text{Size})^{0.39}$

2. Fluff RDF facility:

   Unit Capital Costs = 161,880 $(315/\text{Size})^{0.39}$

3. Total Capital Costs = Unit Capital Costs * TPD

Annualized Costs[c]

1. Operating and Maintenance Costs excluding waste disposal:

   Costs = (12.5 - 0.00115 TPD) * Total Capital Costs/100

2. Capital Recovery[c]:

   Costs = CRF * Total Capital Costs

3. Waste Disposal of Bottom Ash:

   $$\text{Costs} = \frac{1}{24} * \left(\frac{100 - \text{WR}}{100}\right) * \text{TPD} * \text{HRS} * \text{WDC}$$

---

[a] Costs are estimated in December 1987 dollars.

[b] Size = combustor RDF feed rate, tons/day
TPD = plant MSW feed rate, tons/day
CRF = capital recovery factor, 0.1315 based on 10 percent interest rate and 15-year economic life
WR = weight reduction of MSW in the combustor, percent
HRS = hours of operation
WDC = waste disposal cost rate, dollars per ton (typically $25/ton)

[c] Applies only to new plants. Capital recovery costs are not estimated for retrofit applications since the capital costs are sunk.

TABLE B-4. PROCEDURE FOR ESTIMATING CAPITAL COSTS FOR NEW FBC'S
(December 1987 dollars)

Total Direct and Indirect Costs:[a]

$$\text{Costs, } 10^3\$ = 64{,}900 * \text{TPD} * (900/\text{TPD})^{0.39}$$

Process Contingency: 20% of total direct and indirect costs

Total Capital FBC Costs: Total direct and indirect costs + process contingency

---

[a] TPD = plant municipal waste feed rate, tons/day.

TABLE B-5. PROCEDURE FOR ESTIMATING ANNUAL OPERATING COSTS FOR FBC'S (December 1987 dollars)

Combustor and Balance of Plant (excludes coarse RDF processing area):

Operating labor (based on 10 man-years, 40 hours/week, $12/hr):

$OL = 10 * 40 * 52 * 12 * (TPD/900) = 277.3 * TPD$

Supervision (based on 3 man-years/year, 40 hours/week, 30% wage rate premium over the operating labor wage):

$SPRV = 3 * 40 * 52 * 12 * 1.3 * (TPD/900) = 108.2 * TPD$

Maintenance labor (based on 3 man-years/year, 40 hours/week, 10% wage rate premium over the operating labor wage):

$ML = 3 * 40 * 52 * 12 * 1.1 * (TPD/900) = 91.5 * TPD$

Maintenance materials: 3% of the total capital costs

Electricity (based on 3 MW power consumption, and electricity rate of $0.046/kwh):

$ELEC = 0.153 * TPD * HRS$

Limestone (based on $40/ton for limestone):

$LIMESTONE = 0.02 * LFEED * HRS * N$

Water (based on 3% blowdown rate and $0.05/1,000 gal):

$WC = 1.86 \times 10^{-6} * STM * HRS$

Waste disposal (based on tipping fee of $25/hr):

$AD = 1.25 \times 10^{-2} * N * HRS * WDR$

Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of maintenance materials costs

Continued

TABLE B-5. (CONCLUDED). PROCEDURE FOR ESTIMATING ANNUAL OPERATING COSTS FOR FBC'S (December 1987 dollars)

Coarse RDF Processing Area:

Total Operating and Maintenance Costs (TOT O&M):

$\text{TOT O\&M} = 4.4 \times 10^{-4} * (12.5 - 0.00115 * \text{TPD}) * \text{TDI}$

Taxes, Insurance, and Administrative Charges:

4% of the total capital cost

Capital Recovery (based on 15 year life and 10% interest rate):

13.15% of the total capital cost

---

[a] OL = operating labor costs, $/yr
SPRV = supervision costs, $/yr
ML = maintenance labor costs, $/yr
ELEC = electricity costs, $/yr
HRS = hours of operation per year
LIMESTONE = limestone costs, $/yr
LFEED = limestone feed rate per unit, lb/hr
N = number of combustors
WC = water costs, $/yr
STM = plant steam production, lb/hr
AD = waste disposal costs, $/yr
WDR = waste disposal rate per unit (bottom and fly ash collected), lb/hr
TPD = plant municipal waste feed rate, tons/day
TDI = total direct and indirect capital costs for FBC plant, $

TABLE B-6. PROCEDURES FOR ESTIMATING CAPITAL COSTS FOR ELECTROSTATIC PRECIPITATORS (ESP'S)[a,b]

Design Equation for Mass-burn and RDF Facilities:

$$SCA = -189.29 \ln \frac{(100 - PMEFF)}{101.89}$$

Design Equation for Modular Units:

- o Use above design equation for large modular units whose flue gas flowrate (Q) is greater than or equal to 30,000 acfm
- o For small modular units whose $Q < 30,000$

$$SCA = -285.7 \ln \frac{(100 - PMEFF)}{79.6}$$

Purchased Equipment Costs

ESP for Massburn and RDF plants and large modular plants[c]:
Costs, 10^3 \$ = $(305.2 + 0.00738 * TPA) * RF * N$

ESP for small modular plants ($Q < 30,000$)[c]:
Costs, 10^3 \$ = $1.08 * (96.3 + 0.015 * TPA) RF * N$

ESP Rebuilds:
Costs, 10^3 \$ = 0.42 * purchased equipment costs for a new ESP (RF = 1)

Ductwork:
Costs, 10^3 \$ = $0.7964 * N * RF * Q^{0.5}$

Fan:
Costs, 10^3 \$ = $1.077 * N * RF * Q^{0.96}$

Installation Direct Costs

= 67% of purchased equipment costs for new ESP and ESP upgrades (i.e., addition of new plate area in existing ESP)
= 33% of purchased equipment costs for ESP rebuilds only

(continued)

TABLE B-6. (Continued)

Indirect Costs

= 54% of purchased equipment costs for mass-burn, RDF, and large modular units with new ESP and ESP upgrades
= $14,000 for small modular units with new ESP and ESP upgrades
= 24% of the purchased equipment costs for ESP rebuilds

Contingency

= 3% of purchased equipment costs

Total Capital Costs

= Purchased equipment costs + installation direct costs + indirect costs + contingency costs

---

[a] Costs are estimated in December 1987 dollars.

[b] PMEFF = particulate matter removal efficiency, percent
SCA = specific collection area, ft^2/1,000 acfm
Q = 125 percent of the actual flue gas flowrate per ESP unit, acfm
TPA = total plate area, ft^2
RF = retrofit factor obtained from Table B-16
N = number of ESP units
L = duct length, feet

[c] Includes taxes and freight of eight percent of the ESP equipment costs. For retrofit applications requiring additional plate area of the existing ESP, TPA is the increase in the plate area.

TABLE B-7. PROCEDURE FOR ESTIMATING ANNUAL OPERATING COSTS FOR ESP's[a,b]

Operating Labor[c] (Based on 1 manhour/shift, labor wage of \$12/hr): OL = 1.5 * N * HRS

Supervision[c]: 15% of the operating labor cost (OL)

Maintenance Labor[c] (Based on 0.5 manhour/shift, 10% wage rate premium over the operating labor wage): ML = 0.825 * N * HRS

Maintenance Materials: 1% of the direct capital costs[d]

Electricity for I.D. Fan (Based on 0.5 inch pressure drop W.C. and electricity rate of \$0.046/kwh): $\text{FANELEC} = 4.50 \times 10^{-6} * \text{FLW} * \text{N} * \text{HRS}$

Electricity Consumed by ESP (Based on 1.5 watts/ft^2 collection area and electricity rate of \$0.046/kwh): $\text{ESPELEC} = 6.9 \times 10^{-8} * \text{TPA} * \text{N} * \text{HRS}$

Ash Disposal (Based on tipping fee of \$25/ton): $\text{AD} = 1.25 \times 10^{-2} * \text{N} * \text{HRS} * \text{WDR}$

Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of maintenance materials costs

Taxes, Insurance, and Administrative Charges: 4% of the total capital costs

Capital Recovery (Based on 15 year life and 10% interest rate): 13.15% of the total capital costs

---

[a] For ESP rebuilds, the only annual costs are for an increase in waste disposal and capital recovery. All costs are estimated in December 1987 dollars.

[b] OL = operating labor costs, \$/yr
N = number of ESP units
HRS = hours of operation
ML = maintenance labor costs, \$/yr
FANELEC = electricity costs for I.D. fan, \$/yr
ESPELEC = electricity costs for ESP, \$/yr
FLW = actual flue gas flowrate per unit, acfm
AD = ash disposal costs, \$/yr
WDR = waste disposal rate per unit, lb/hr

[c] Labor costs are estimated only for new ESP's. The labor costs for ESP retrofits (including upgrades and rebuilds) are zero.

[d] Direct capital costs are the sum of purchase equipment costs and installation direct costs.

TABLE B-8. PROCEDURES FOR ESTIMATING CAPITAL COSTS FOR DRY SORBENT INJECTION[a,b]

Purchased Equipment Costs, 10^3 \$

1. Lime Storage Silo with Vibrator, Baghouse, and Flow Control Value (Based on 15-day lime supply)

   - o For storage volumes (V) less than or equal to 2,300 ft^3 (one storage silo per plant),
     Costs = 1.05 * (34.2 + 0.016V) * RF
   - o For storage volumes between 2,300 and 4,600 ft^3 (two storage silos per plant),
     Costs = 2.10 * (34.2 + 0.016V) * RF
   - o For storage volume greater than 4,600 ft^3 (two storage silos per plant),
     Costs = 2.10 * (63 + 0.0038V) * RF

2. Feed Bins

   - o For duct sorbent injection (one feed bin per combustor),
     Costs = 0.0906 * N * RF * (SF)$^{0.6145}$
   - o For furnace sorbent injection (two feed bins per combustor),
     Costs = 0.1812 * N * RF * (SF)$^{0.6145}$

3. Gravimetric Feeders

   - o For duct sorbent injection (one feeder per combustor),
     Costs = 1.024 * (0.000289 SF + 9.293) * N * RF
   - o For furnace sorbent injection (two feeders per combustor),
     Costs = 2.048 * (0.000289 SF + 9.293) * N * RF

4. Pneumatic Conveyor (Based on 400 Feet Length)

   Costs = 1.05 * (26.4 + 0.0073 SF + 0.4 * [12.8 + 11.23 SF$^{0.23}$]) N * RF

5. Injection Ports

   - o For duct sorbent injection (one injection port per combustor),
     Costs = 1.05 * (22.2 + 0.0014 SF) * N * RF
   - o For furnace sorbent injection (two injection ports per combustor),
     Costs = 2.10 * (22.2 + 0.0014 SF) * N * RF

(continued)

TABLE B-8. (Continued)

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6. Reactor Vessel (optional for duct sorbent injection to increase flue gas and sorbent contact): Costs = $34 * (Q * 1.25/6{,}150)^{0.64}$

7. Fabric Filter[c]

   Costs = $0.1482 * N * RF * Q^{0.7043}$

8. Induced Draft Fan

   Costs = $(1.167 * N * RF * Q^{0.96})/1{,}000$

9. Ductwork

   Costs = $(0.8627 * N * RF * L * Q^{0.5})/1{,}000$

Installation Direct Costs

= 30% of dry sorbent injection equipment costs + 72% of fabric filter and auxiliary equipment costs

Indirect Costs

= 33% of direct costs (equipment + installation costs) for dry sorbent injection +
42% of equipment cost for the fabric filter and auxiliary equipment

Contingency

= 50% of the sum of direct and indirect costs

Total Capital Costs

= Total Direct Costs + Indirect Costs + Contingency Costs

---

[a] All costs are estimated in December 1987 dollars.

[b] SF = lime feed rate per unit, lb/hr
Q = 125 percent of the actual flue gas flowrate to the fabric filter per unit, acfm
V = lime storage silo volume for the plant, ft^3
N = number of units
RF = Retrofit factor. For retrofit applications, use retrofit factor of 1.1 for sorbent injection equipment. Retrofit factors for fabric filter and auxiliary equipment are obtained from Table B-16.
L = duct length, feet

[c] Fabric filters are used for new applications and for retrofit applications where no ESP exists. For plants with existing ESP's, costs for upgrading the ESP are estimated from Table B-6.

TABLE B-9. PROCEDURES FOR ESTIMATING ANNUAL OPERATING COSTS FOR DRY SORBENT INJECTION[a,b]

Operating Labor (Based on 2 manhours/shift, wage of $12/hr): CL = 3.0 * N * HRS

Supervision: 15% of the operating labor cost (OL)

Maintenance Labor: (Based on 0.5 manhour/shift, 10% wage rate premium over the operating labor wage): ML = 0.825 * N * HRS

Maintenance Materials: 5% of the direct capital costs of the sorbent injection equipment[c]

Electricity (Based on electricity rate of $0.046/kwh): ELEC = 5.25 x 10^{-6} * (251,850 + 52.56 * SF) * N * HRS

Lime: (Based on $80/ton for hydrated lime and acid gas stoichiometric ratio of 2:1): LIME = 0.04 * SF * N * HRS

Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of maintenance materials costs

Taxes, Insurance, and Administrative Charges: 4% of the total capital costs

Capital Recovery (Based on 15 year life and 10% interest rate): 13.15% of the total capital costs

---

[a] All costs are estimated in December 1987 dollars.

[b] OL = operating labor costs, $/yr
N = number of units
HRS = hours of operation
ML = maintenance labor costs, $/yr
ELEC = dry sorbent injection electricity costs, $/yr
SF = lime feed rate per unit, lb/hr
LIME = lime costs, $/yr

[c] Direct capital costs are the sum of purchase equipment costs and installation direct costs.

TABLE B-10. PROCEDURES FOR ESTIMATING ANNUAL OPERATING COSTS FOR FABRIC FILTERS[a,b]

Operating Labor (Based on 2 manhours/shift, wage of \$12/hr): OL = 3.0 * N * HRS

Supervision: 15% of the operating labor cost (OL)

Maintenance Labor: (Based on 1 manhour/shift, 10% wage rate premium over the operating labor wage): ML = 1.65 * N * HRS

Maintenance Materials: 5% of the direct capital costs of both the fabric filter and auxiliary equipment[c]

Bag Replacement: (Based on \$1.35/ft^2, gross air-to-cloth ratio of 3:1, 2 year bag life, and 10% interest rate): BAG = 0.2593 * Q * N

Electricity (Based on 12.5 inches W.C. pressure drop and electricity rate of \$0.046/kwh): ELEC = 1.12×10^{-4} * FLW * N * HRS

Compressed Air (Based on 2 scfm of air/1,000 acfm flue gas and \$0.17/1,000 scfm for compressed air): AIR = 2.01×10^{-5} * FLW * N * HRS

Ash Disposal (Based on tipping fee of \$25/ton): AD = 1.25×10^{-2} * N * HRS * WDR

Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of maintenance materials costs

Taxes, Insurance, and Administrative Charges: 4% of the total capital costs

Capital Recovery (Based on 15 year life and 10% interest rate): 13.15% of the total capital costs

---

[a] All costs are estimated in December 1987 dollars.

[b] OL = operating labor costs, \$/yr
N = number of units
HRS = hours of operation
ML = maintenance labor costs, \$/yr
BAG = bag replacement costs, \$/yr
Q = 125 percent of the inlet flue gas flowrate per unit, acfm
FLW = actual flue gas flowrate per unit, acfm
ELEC = electricity costs, \$/yr
AIR = compressed air costs, \$/yr
WDR = ash disposal rate from fabric filter per unit, lb/hr

[c] Direct capital costs are the sum of the purchase equipment costs and installation direct costs.

TABLE B-11. PROCEDURES FOR ESTIMATING CAPITAL COSTS OF STAND-ALONE SPRAY DRYER AND SPRAY DRYER/FABRIC FILTERS[a,b]

Direct Costs

SD/FF Unit[c]: Costs, 10^3 \$ = $8.053 * N * RF * (Q)^{0.517}$

Stand-Alone SD Unit: Costs, 10^3 \$ = $8.428 * N * RF * (Q)^{0.460}$

Ductwork[c]: Costs, 10^3 \$ = $(1.3868 * N * RF * L * Q^{0.5})/1{,}000$

Fan[c]: Costs, 10^3 \$ = $(1.8754 * N * RF * Q^{0.96})/1{,}000$

Indirect Costs = 33% of direct costs

Contingency = 20% of sum of direct and indirect costs

Total Capital Investment = Direct Costs + Indirect Costs + Contingency Costs

---

[a] All costs are estimated in December 1987 dollars.

[b] Q = 125% of the actual flue gas flowrate, acfm
N = number of units
RF = retrofit factor obtained from Table B-16
L = Duct length per unit, feet

[c] The total installed costs are assumed to be 133 percent of the direct capital costs.

TABLE B-12. PROCEDURE FOR ESTIMATING ANNUAL OPERATING COSTS FOR STAND-ALONE SPRAY DRYERS AND SPRAY DRYER/FABRIC FILTERS[a,b]

Operating Labor for SD/FF (Based on 4 manhours/shift, labor wage of $12/hr): OL = 6.0 * N * HRS

Operating Labor for SD (Based on 2 manhours/shift, labor wage of $12/hr): OL = 3.0 * N * HRS

Supervision: 15% of the operating labor cost (OL)

Maintenance Labor for SD/FF: (Based on 2 manhours/shift, 10% wage rate premium over the operating labor wage): ML = 3.3 * N * HRS

Maintenance Labor for SD: (Based on 1 manhour/shift, 10% wage rate premium over the operating labor wage): ML = 1.7 * N * HRS

Maintenance Materials: 2% of the total direct capital costs

Bag Replacement for SD/FF: (Based on $\$1.35/\text{ft}^2$, gross air-to-cloth ratio of 3:1, 2 year bag life, and 10% interest rate): BAG = 0.2593 * Q * N

Electricity for I.D. Fan for SD/FF (Based on 12.5 inches W.C. pressure drop and electricity rate of $0.046/kwh): FANELEC = 1.12 x 10^{-4} * FLW * N * HRS

Electricity for I.D. Fan for SD (Based on 5.5 inches W.C. pressure drop and electricity rate of $0.046/kwh): FANELEC = 4.93 x 10^{-5} * FLW * N * HRS

Electricity for Atomizer (Based on 6 kw per 1,000 lb/hr of slurry feed + 15 kw and electricity rate of $0.046/kwh): ATELEC = 0.046 * (0.006 * (SF + WTR) + 15) * N * HRS

Electricity for Pump (Based on 20 feet of pumping height, 10 psi discharge pressure, 10 ft/sec velocity in piping): PUMPELEC = 1.291 X 10^{-6} * (SF + WTR) * N * HRS

Compressed Air for SD/FF (Based on 2 scfm of air/1,000 acfm flue gas and $0.17/1,000 scfm for compressed air): AIR = 2.01 x 10^{-5} * FLW * N * HRS

Water: (Based on Flue gas cooling to 300°F and $0.50/1,000 gal): WC = 6.00 x 10^{-5} * WTR * N * HRS

Lime: (Based on $70/ton for quick lime): LIME = 0.035 * SF * N * HRS

TABLE B-12. (Continued)

Ash Disposal for SD/FF (Based on tipping fee of $25/ton): $AD = 1.25 \times 10^{-2} * N * HRS * WDR$

Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of maintenance materials costs

Taxes, Insurance, and Administrative Charges: 4% of the total capital costs

Capital Recovery (Based on 15 year life and 10% interest rate): 13.15% of the total capital costs

---

[a] All costs are estimated in December 1987 dollars.

[b] OL = operating labor costs, $/yr
N = number of units
HRS = hours of operation
ML = maintenance labor costs, $/yr
BAG = bag replacement costs, $/yr
Q = 125 percent of the actual flue gas flowrate to the fabric filter per unit, acfm
FLW = actual flue gas flowrate to the fabric filter per unit, acfm
SF = lime feed rate per unit, lb/hr
WTR = water rate per unit, lb/hr
FANELEC = electricity costs for the fan, $/yr
ATELEC = electricity costs for the atomizer, $/yr
PUMPELEC = electricity costs for the the pumps, $/yr
AIR = compressed air costs, $/yr
WC = water costs, $/yr
LIME = lime costs, $/yr
AD = ash disposal costs, $/yr
WDR = waste disposal rate, lb/hr

TABLE B-13. PROCEDURES FOR ESTIMATING CAPITAL COSTS FOR HUMIDIFICATION[a,b]

Purchased Equipment Costs, 10^3 $

1. Humidification Chamber and Pumps:

   Costs = (0.438 * Q + 80,220) N * RF/1,000

2. Ductwork:

   Costs = $(1.16 * L * Q^{0.5}) * N * RF/1,000$

Installation Direct Costs = 56% of Purchase Equipment Costs

Indirect Costs = 32% of Purchase Equipment Costs

Contingency = 3% of the Purchase Equipment Costs

Total Capital Costs = Purchased Equipment Costs + Installation Direct Costs + Indirect Costs

= 191% of Purchase Equipment Costs

---

[a] All costs are estimated in December 1987 dollars.

[b] Q = 125% of the actual flue gas flowrate, acfm
L = duct length per unit, feet
N = number of units
RF = retrofit factor obtained from Table B-16

TABLE B-14. PROCEDURE FOR ESTIMATING ANNUAL OPERATING COSTS FOR HUMIDIFICATION[a,b]

Operating Labor (Based on 0.5 manhours/shift, labor wage of $12/hr): OL = 0.75 * N * HRS

Supervision: 15% of the operating labor cost (OL)

Maintenance Labor: (Based on 0.5 manhour/shift, 10% wage premium over the operating labor wage): ML = 0.825 * N * HRS

Maintenance Materials: 1% of the total capital costs

Electricity for Pumps[c] (Based on 20 feet of pumping height, 100 psi discharge pressure, 10 ft/sec velocity in piping, and electrical rate of $0.046/kwh): ELEC = 7.3 X 10^{-6} * N * HRS * WTR

Water (Based on $0.50/1,000 gal)[c]: WC = 6.0 x 10^{-5} * WTR * N * HRS

Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of maintenance materials costs

Taxes, Insurance, and Administrative Charges: 4% of the total capital costs

Capital Recovery (Based on 15 year life and 10% interest rate): 13.15% of the total capital costs

---

[a] All costs are estimated in December 1987 dollars.

[b] OL = operating labor costs, $/yr
N = number of units
HRS = hours of operation
ML = maintenance labor costs, $/yr
WTR = water injection rate, lb/hr
ELEC = electricity costs, $/yr
WC = water costs, $/yr

[c] WTR = (Tin-Tout) * Qs * (1-MOIST/100)/940

where
Tin = inlet flue gas temperature, °F
Tout = outlet flue gas temperature, °F
Qs = flue gas flow rate per unit, scfm
MOIST = moisture content in flue gas, volume percent

TABLE B-15. CONTINUOUS MONITORING COST SUMMARY[a]

Pollutant compliance options	Method	Capital costs ($1,000)	Operating costs ($1,000/yr)[c]	Annualized costs ($1,000/yr)[d]
PM only	Opacity[b]	61	8	16
Acid gas only	SO_2 (inlet and outlet)	67	10	19
	HCl (inlet and outlet)	140	74	92
	O_2/CO_2	19	15	18
	Data Reduction System	31	4	8
	Total	256	103	137
PM + acid gas	Opacity[b]	61	8	16
	SO_2 (inlet and outlet)	67	10	19
	HCl (inlet and outlet)	140	74	92
	O_2/CO_2	19	15	18
	Total	286	107	145

[a] All costs are reported in December 1987 dollars. For multiple control units, multiply these costs by the number of units.

[b] Includes costs for automatic data reduction system.

[c] Based on 2 certifications/year and maintenance requirements of 0.5 man-hour/day for both opacity and O_2/CO_2 monitors and 1 man-hour/day for SO_2 and HCl monitors.

[d] Annualized costs include annual operating costs and capital charges on equipment and installation costs. Capital charges are based on a 15-year equipment life at 10 percent interest rate.

TABLE B-16. SITE ACCESS AND CONGESTION FACTORS FOR RETROFITTING APCD EQUIPMENT AT EXISTING PLANTS

Retrofit factor	Congestion level	Guidelines for selecting retrofit factor
1.02	Base case	Interferences similar to a new plant with adequate crew work space. Free access for cranes. Area around combustor and stack adequate for standard layout of equipment.
1.08	Low	Some aboveground interferences and work space limitations. Access for cranes limited to two sides. Equipment cannot be laid out in standard design. Some equipment must be elevated or located remotely.
1.25	Medium	Limited space. Interference with existing structures or equipment which cannot be relocated. Special designs are necessary. Crane access limited to one side. Majority of equipment elevated or remotely located.
1.42	High	Severely limited space and access. Crowded working conditions. Access for cranes blocked from all sides.

TABLE B-17. PROCEDURE FOR ESTIMATING SCOPE ADDER CAPITAL COSTS[a,b]

Direct and Indirect Costs

1. New Ducting:

   Costs, 10^3 \$ = $1.844 * L * N * RF * Q^{0.5}$

2. New I.D. Fan:

   Costs, 10^3 \$ = $2.493 * N * RF * Q^{0.96}$

3. New Stacks (costs per stack):

   o For a lined acid resistant stack,

   Costs, 10^3 \$ = $[26.2 + 0.089 * H * (1 + 4.14 * D)]$ for $D \geq 5$

   Costs, 10^3 \$ = $[26.2 + 0.080 * H * (1 + 4.33 * D)]$ for $D < 5$

   o For a unlined stack,

   Costs, 10^3 \$ = $[26.2 + 0.0625 * H * (1 + 2.59 * D)]$ for $D \geq 5$

   Costs, 10^3 \$ = $[26.2 + 0.087 * H * (1 + 2.2 * D)]$ for $D < 5$

Contingency = 20% of the direct and indirect costs

Total Capital Costs = Direct Costs + Indirect Costs + Contingency Costs

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[a] All costs are estimated in December 1987 dollars.

[b] L = duct length per unit, feet
N = number of units
RF = retrofit factor obtained from Table B-16
Q = 125% of the actual flue gas flowrate, acfm
H = stack height, feet
D = stack diameter, feet

TABLE B-18. DOWNTIME COST PROCEDURE[a]

Capital Costs[b]

Loss of Tipping Fees:

$$\text{Costs, \$} = \left(\frac{WR}{100}\right) * \left(\frac{TPD}{24}\right) * WDC * HRS * (DT/12)$$

Loss of Energy (Steam or Electricity):

$$\text{Costs, \$} = \frac{TPD}{24} * ER * HRS * DC * (DT/12)$$

Annualized Costs (Capital Recovery)[c]

$$\text{Costs, \$} = CRF * \text{Downtime Capital Costs}$$

---

[a] Costs are estimated in December 1987 dollars. Apply only to retrofit applications.

[b] WR = weight reduction of waste in the combustor, percent
WDC = waste disposal cost rate, dollars per ton (typically $25/ton)
TPD = plant waste feed rate, tons/day
HRS = hours of operation
DT = downtime, months
ER = energy cost rate, dollars per ton ($24.84/ton for mass burn waterwall units, $36.16/ton for RDF units, and $19.52/ton for modular and mass burn refractory units with heat recovery)

[c] CRF = capital recovery factor, 0.1315 based on 10 percent interest rate and 15-year economic life

2.0 PROCEDURES FOR NEW PLANTS

This section presents procedures for estimating costs of new MWC plants. The capital and annualized costs of a new plant include the combustors and associated equipment, air pollution control devices (APCD's), and continuous emission monitoring equipment. Section 2.1 presents the procedures for costing the MWC combustors and associated equipment (denoted as the balance of plant). Procedures for costing electrostatic precipitators (ESP's), dry sorbent injection (DSI), and spray drying (SD) are presented in Sections 2.2, 2.3, and 2.4, respectively. Section 2.5 presents compliance monitoring equipment costs for opacity, HCl, SO_2, O_2, and CO_2.

2.1 COMBUSTORS AND BALANCE OF PLANT

This section presents procedures for estimating the combustor and associated equipment costs, excluding the air pollution control devices (APCD's), for four types of combustors: modular, mass burn, refuse-derived fuel (RDF), and fluidized bed combustion (FBC). The capital cost procedures for each combustor type with the exception of fluidized bed combustors were developed from data presented in Frost and Sullivan.[1] The operating cost procedures were developed from responses to an information request which EPA sent to MWC operators under authority of Section 114 of the Clean Air Act. Capital and operating cost procedures for FBC's were developed from vendor data.

Sections 2.1.1, 2.1.2, 2.1.3, and 2.1.4 present the cost procedures for modular, mass burn, RDF, and FBC facilities, respectively. Detailed cost data suitable for direct estimation of capital costs for MWC's of various types and sizes were not available. Therefore, the procedures presented in these sections are based on scaling the capital cost of typical size facilities of each combustor type. The capital costs estimated for these facilities were all based on Frost and Sullivan (with the exception of FBC's) to minimize inconsistencies in cost assumptions among combustor types. To facilitate use of these costs with those presented in subsequent sections of this report, the original combustor capital cost estimates were revised to exclude the cost of the APCD. Indirect cost estimations for general facilities and engineering fees are also based on Frost and Sullivan.

Capital costs were updated from 1985 dollars in Frost and Sullivan to December 1987 dollars using the Chemical Engineering Plant Cost Index.

2.1.1 Modular Units

2.1.1.1 Overview of Technology. Modular combustors are prefabricated units which are generally used to combust unprocessed MSW. Individual combustor units typically range in size from 5 to 150 tons per day (tpd).

Modular combustors are of two general designs, starved-air and excess-air. In typical starved-air combustors, MSW is ram-fed into a primary combustion chamber with a moving grate. Primary air is fed up through the grate at substoichiometric conditions. Volatile gases released from the heated MSW enter a secondary chamber where sufficient air and supplemental fuel, typically natural gas or oil, are supplied to complete combustion. Excess-air combustors provide air in excess of stoichiometric requirements in the primary combustion chamber. Additional air and supplemental fuel may be added in a secondary chamber to assure complete combustion.

Modular combustors which recover energy typically do so in waste heat boilers following the combustion chambers. The steam produced can be sold directly to users or used to generate electricity.

2.1.1.2 Capital Cost Procedures. The original modular combustors were relatively simple in design and had low capital costs. As a result of subsequent design changes, the capital costs of modular units have increased but are still relatively low compared to other MWC technologies on a "ton per day of MSW capacity" basis. However, thermal efficiencies are also lower, decreasing the cost advantage for facilities when designed primarily for heat recovery. Because of their modular design, little or no economy of scale exists as a function of facility size.[2]

As shown in Table 2.1-1, capital costs were obtained for a 50 tpd modular facility without heat recovery and for a 100 tpd facility consisting of two modular starved-air combustors, one waste heat boiler (17,000 lb/hr steam), and an optional 1.475 MW steam turbine. Costs for excess air modular units are expected to be similar. Based on the costs presented in this

TABLE 2.1-1. CAPITAL COSTS FOR MODULAR MWC's[a]

Equipment	Costs, $1,000's[b]
A. Without Heat Recovery	
Combustors (1 @ 50 tpd)	1,125
Total[c]	1,125
Engineering Fees	88
Total Capital Costs	1,210
Unit Capital Costs	$24,300 per tpd of MSW processed
B. With Electricity Generation (Without Electricity Generation)[d]	
Combustors (2 @ 50 tpd)	2,250
Waste Heat Boiler (1 @ 17,000 lb/hr)	770
Turbine/Generator (1 @ 1,475 kW)	2,050 (0)
Total[c]	5,070 (3,020)
Engineering Fees	390 (232)
Total Capital Costs	5,460 (3,250)
Unit Capital Costs	$54,600 ($32,500) per tpd of MSW processed

[a] Reference 1, p. 128.

[b] December 1987 dollars.

[c] Represents total direct costs (sum of the equipment and installation costs).

[d] Costs in parenthesis represent modular MWC's producing steam without generating electricity.

table, the following procedure can be used to estimate capital costs for modular facilities:

- Modular facilities without heat recovery = \$24,300 per tpd capacity
- Modular facilities producing steam = \$32,500 per tpd capacity
- Modular facilities generating electricity = \$54,600 per tpd capacity

The heat recovery boiler and turbine/generator set effectively double the cost of the facility.

Because modular combustors are packaged units which require little site preparation or extra equipment, no separate costs were included for general facilities (foundations and building) or MSW or ash handling systems.

2.1.1.3 Operating Cost Procedures. Annual operating cost procedures were developed from analysis of cost data provided by five plants with modular combustors in their responses to the Section 114 questionnaire. All operating costs except for capital recovery and waste disposal costs were examined. The approach used in developing the operating cost procedures was to correlate the ratio of the operating to capital costs with facility capacity (tpd) and annual operating hours. The following best fit equation was derived for these facilities, all of which were rated below 150 tpd and were operating less than 6,000 hours per year:

$$\text{Ratio} = 10 - 0.23\ \text{tpd} + 0.006\ \text{hrs} \qquad (1)$$

where, Ratio = percentage of operating to capital plant costs
tpd = facility waste feed rate, tons/day
hrs = annual operating hours.

For modular facilities outside these size or operating hour limits, the following equation can be used to estimate annual operating cost based on mass burn refractory wall facilities as discussed in Section 2.1.2.3:

$$\text{Ratio} = 15.7 - 0.00115\ \text{tpd} \qquad (2)$$

It is assumed that the ratio of operating to capital costs for modular and mass burn refractory wall facilities are similar, since the design and equipment arrangements of both MWC types are similar (i.e., use of waste heat boilers for heat recovery).

Capital recovery costs are calculated as 13.15 percent of the total capital costs, based on a 10 percent interest rate and 15 year economic life. Waste disposal costs are estimated by the following equation based on a \$25/ton landfill tipping fee:

$$WDC = \$25 * \frac{100 - RED}{(100)} * TPD * \frac{(HRS)}{(24)} \quad (3)$$

where, WDC = waste disposal costs, \$/yr

RED = weight reduction of MSW within the combustor, percent

TPD = facility waste feed rate, tons/day

HRS = annual operating hours.

Therefore, the total annualized costs are obtained by summing the annual operating costs calculated from Equations 1 or 2, the capital recovery costs, and the waste disposal costs estimated from Equation 3. All costs are presented in December 1987 dollars.

2.1.2 Mass Burn Units

2.1.2.1 Overview of Technology. Mass burn combustors are field-erected units used to combust unprocessed municipal solid waste. Mass burn combustors range in unit size from 50 to 1,000 tons/day for a combustor unit and from 50 to several thousand tons/day for a facility.

In a mass burn combustor, municipal waste is gravity- or ram-fed to a single combustion chamber. Several different grate designs can be used to move the waste through the combustion chamber. Air is supplied in excess of stoichiometric requirements through the grates (underfire air) and into the combustion chamber above the grates (overfire air).

Either waterwall or waste heat boilers are used for recovering heat. In units with waterwalls, boiler tubes are built into the walls of the combustion chamber. Additional heat recovery sections can include superheaters, economizers, and air preheaters. In units with waste heat boilers, the combustion chamber is refractory lined and steam is generated downstream of the combustion chamber.

Electricity is generally produced on-site from steam. Some facilities sell both electricity and steam. Because of the large quantity of steam produced by large mass burn units, production and sale of only steam is unlikely unless a very large industrial facility with a consistent steam demand is nearby.

Waterwall units are typically larger (100 to 1,000 tons/day per unit) than units with waste heat boilers (50 to 375 tons/day per unit). Most new units are of waterwall design. No units are projected to be built without heat recovery.

2.1.2.2 Capital Cost Procedures. As shown in Table 2.1-2, the estimated capital cost for a mass burn facility is $60,700 per tpd capacity. This cost is based on an 860 tpd facility consisting of two 430 tpd combustors with 174,000 lb/hr of steam capacity and a 20 MW turbine. For the same facility without a steam turbine, the estimated capital costs are $50,420 per tpd capacity. The capital costs for waterwall units and refractory units with waste heat boilers are assumed to be roughly equal.

To account for the economy of scale of mass burn facilities, capital costs reported by Frost and Sullivan were correlated with facility size to yield the following scaling equation:[3]

$$C = 60{,}700\,(430/\text{size})^{0.39}, \quad \text{with electrical generation} \qquad (4)$$

$$C = 50{,}420\,(430/\text{size})^{0.39}, \quad \text{without electrical generation} \qquad (5)$$

where, C = new facility capital costs in December 1987 dollars per tpd

size = size per combustor in tpd.

Capital cost for a combustor is estimated using equations 4 or 5 to calculate cost on a dollar per tpd basis and then multiplying this cost by plant capacity in tons per day.

2.1.2.3 Operating Cost Procedures. Annual operating cost procedures were developed from analysis of cost data provided by six plants (4 of which were mass burn waterwall and 2 were RDF facilities) in their responses to the Section 114 questionnaire. The RDF facility operating cost data were

TABLE 2.1-2. CAPITAL COSTS FOR A 860 TON/DAY MASS BURN MWC WITH (WITHOUT) ELECTRICITY GENERATION[a,b]

Equipment	Costs, $1,000[c]					
	Purchase Cost		Installation Cost		Total Installed Cost	
Waterwall Combustors (2 @ 430 ton/day)	6,742		3,519		10,261	
Refuse Cranes (2)	1,245		498		1,743	
Weight Scales (2)	614		302		916	
Fans and Ducts (2)	1,245		368		1,613	
Ash Handling System	2,933		440		3,373	
Dump Condenser (1)	778		547		1,325	
Stacks (2)	880		365		1,245	
Water Supply and Treatment	1,611		189		1,800	
Piping System	3,127		975		4,102	
Electrical	1,562		293		1,855	
Instruments and Controls	2,073		195		2,268	
Insulation and Paint	519		293		812	
Cooling Tower (1)	3,105		440		3,545	
Turbine/Generator (1 @ 20 MW)	6,318	(0)	1,100	(0)	7,418	(0)
Total Capital Costs	32,752	(26,434)	9,524	(8,424)	42,276[d]	(34,858)[d]
General Facilities						
Foundations	2,544	(2,053)	1,664	(1,472)	4,208	(3,525)
Building and Structural	1,487	(1,200)	172	(152)	1,659	(1,352)
Total General Facilities	4,031	(3,253)	1,836	(1,624)	5,867	(4,877)
Engineering Fees	--		4,097	(3,624)	4,097	(3,624)
Total Capital Costs	36,783	(29,687)	15,457	(13,672)	52,240	(43,359)
Unit Capital Costs					$60,700	(50,420) per tpd of MSW processed

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[a] Reference 1, p. 113.

[b] Costs in parentheses represent mass burn MWC's producing steam without generating electricity.

[c] December 1987 dollars.

[d] Represents total direct costs (sum of equipment and installation costs).

analyzed along with the mass burn cost data because of limited data for mass burn plants and because both combustor types are constructed similarly (i.e., field erected) and will likely be operated for a similar number of hours per year. All operating costs except for capital recovery and waste disposal costs were examined. The approach used in developing the operating cost procedures was to correlate the ratio of the operating to capital costs with facility capacity (tpd) and operating hours. The following best-fit equation was derived for mass-burn waterwall facilities:

$$\text{Ratio} = 12.5 - 0.00115\ \text{tpd} \qquad (6)$$

where, Ratio = percentage of operating to capital plant costs
tpd = facility waste feed rate, tons/day.

For mass burn refractory wall facilities, no equation could be derived due to the limited amount of cost data. However, by comparing the Section 114 data for waterwall and refractory wall facilities, the operating to capital cost percentages for refractory facilities were roughly 3 percent higher than those for mass-burn waterwall facilities. Therefore, to estimate the annual operating cost for mass-burn refractory wall facilities, the equation for mass-burn waterwall (Equation 4) was modified to give the following:

$$\text{Ratio} = 15.7 - 0.00115\ \text{tpd}. \qquad (7)$$

Equations 6 and 7 are multiplied by equation 5 to estimate annual operating costs for mass burn waterwall and mass burn refractory wall facilities, respectively.

Capital recovery costs for both types of mass burn facilities are calculated as 13.15 percent of the total capital costs, based on a 10 percent interest rate and 15 year economic life. Waste disposal costs are estimated by equation 3 presented for modular facilities (Section 2.1.1.3). Therefore, the total annualized costs are obtained by summing the annual operating costs calculated from equations 6 or 7, capital recovery costs, and waste disposal costs calculated from equation 3.

2.1.3 RDF Units

2.1.3.1 Overview of Technology. Refuse-derived fuel MWC's are field-erected units used to combust processed MSW. Individual RDF combustors typically combust 180 to 1,200 tons/day of RDF. Plant sizes range from 180 to several thousand tons/day of RDF.

Refuse-derived fuel processing includes removal of noncombustible materials and shredding of the remaining material. The two types of RDF considered in this analysis include coarse RDF (cRDF) suitable for combusting in a specially designed RDF combustor and fluff RDF (fRDF) suitable for suspension firing in a utility or industrial boiler.

Coarse RDF production generally includes primary shredding and ferrous metals recovery. The RDF material is reduced in size to 4 to 6 inches. A weight reduction due to metals recovery of approximately 6 percent is assumed; energy recovery is nearly 100 percent. Fluff RDF production usually includes crushing, initial trommel screening, and magnetic separation followed by primary shredding, air classification, and secondary shredding. This processing removes oversized combustibles and glass as well as nonferrous metals and reduces the size to below 2 inches. The associated weight reduction is approximately 20 percent of the unprocessed MSW; energy recovery is roughly 97 percent. Production of both types of RDF includes a dust control system to prevent fugitive emissions of fine dust generated by RDF processing equipment.

Existing RDF combustion is based almost entirely on use of cRDF in spreader strokers. In general with this technology, cRDF is thrown to the rear of the furnace by a dry-swept stoker. Fine particles are burned in suspension, and heavier materials fall to the grate and are combusted. A traveling grate moves the materials to the front of the furnace during which time combustion is completed. The heat from combustion is recovered using radiant waterwall and convective heat transfer. Electricity is generally generated on-site, especially with the larger units. Both electricity and steam may be sold.

TABLE 2.1-3. CAPITAL COSTS FOR A COARSE RDF FACILITY WITH ELECTRICITY GENERATION (CAPACITY = 850 TONS/DAY MSW, 800 TONS/DAY RDF)[a]

Equipment	Costs, $1,000's[b] Purchase Cost	Installation Cost	Total Installed Cost
Waterwall Combustors (2 @ 400 ton/day)	5,085	2,544	7,629
Front End Loaders (9)	713	--	713
Primary Shredders (2)	1,016	508	1,524
Weigh Scales (2)	639	307	946
Magnetic Separation System (2)	205	68	273
Fans and Ducts (2)	993	273	1,266
Ash Handling System (2)	2,441	368	2,809
Dump Condenser (1)	916	670	1,586
Dust Control System	508	205	713
Stacks (2)	689	303	992
Water Supply and Treatment	1,985	225	2,210
Piping System	3,663	1,221	4,884
Electrical	1,657	375	2,032
Instruments and Controls	2,339	253	2,592
Insulation and Paint	539	307	846
Cooling Tower	3,586	506	4,092
Turbine/Generator (1 @ 24 MW)	7,627	1,345	8,972
Total Capital Costs	34,601	9,478	44,079[c]
General Facilities			
Foundation	3,945	2,704	6,649
Building and Structural	3,283	388	3,671
Total General Facilities	7,228	3,092	10,320
Engineering Fees	--	4,512	4,512
Total Capital Costs	41,829	17,082	58,911
Unit Capital Costs			$73,600 per tpd of RDF

[a] Reference 1, p. 117.

[b] December 1987 dollars.

[c] Represents total direct costs (sum of equipment and installation costs).

Fluff RDF has been co-fired with coal in several existing utility and industrial boilers.[4] Of the total heat input, fRDF generally represents less than 10 percent.

2.1.3.2 Capital Cost Procedures. The cost procedures developed for RDF-fired MWC's are based on cRDF. As shown in Table 2.1-3, the estimated capital cost for a cRDF facility is $73,600 per tpd of cRDF capacity. This cost was based on a cRDF facility designed to process 850 tpd of MSW into 800 tpd of cRDF. The cRDF is combusted in two identical waterwall boilers to generate 203,000 lb/hr of steam for a 24 MW turbine. The major equipment items and associated materials and labor costs for the cRDF production/combustion facility are presented in Table 2.1-3.

Because no data are available from Frost and Sullivan for other cRDF facility sizes, it is assumed that the economy of scale of RDF facilities is the same as mass-burn facilities, since both mass-burn and RDF combustors are field-constructed. Similar indirect installation costs would be incurred for both MWC types. Therefore, the following equation can be used to estimate capital cost for cRDF facilities:

$$C = 73,600 \ (400/\text{size})^{0.39} \qquad (8)$$

where, C = new cRDF facility capital costs in December 1987 dollars per ton of RDF

size = size per combustor in tons RDF/day.

2.1.3.3 Operating Cost Procedures. Annual operating cost procedures were developed from analysis of cost data as discussed in Section 2.1.2.3. To estimate the annual operating costs except for capital recovery and waste disposal costs, equation 6 presented in Section 2.1.2.3 can be used.

Capital recovery costs are calculated as 13.15 percent of the total capital costs, based on a 10 percent interest rate and 15 year economic life. Waste disposal costs are estimated using Equation 3 presented in Section 2.1.1.3. It should be noted that the annual cost procedures do not include estimates on sale of recoverable material such as metal and glass.

2.1.4 FBC Units

2.1.4.1 Overview of Technology. Fluidized-bed combustors are field-erected units used to combust RDF. Only three RDF-fired FBC plants are currently operating in the U. S. Existing and currently planned FBC's are designed to combust 195 to 500 tpd of RDF. Plant sizes range from 195 to 1,200 tpd of RDF.[5] For costing purposes, it is assumed that a cRDF fuel processing facility is included in the design of the FBC facility. Two basic FBC designs exist: bubbling bed and circulating fluidized-bed.

In a bubbling-bed combustor, the RDF burns in a turbulent bed of heated noncombustible material, such as limestone or sand. Typical bed temperatures are from 1,450 to 1,700°F. As with conventional combustors, primary combustion air is introduced underneath the bed, but at a flowrate high enough to suspend or "fluidize" the solid particles in the bed. Secondary combustion air is introduced through ports in the upper part of the combustor to complete the combustion process. If good mixing between air and combustible waste is achieved, the amount of excess air required for complete combustion is similar to conventional RDF combustors. In addition, by adding limestone to the bed, SO_2 and HCl can be removed from the flue gas to reduce acid gas emissions. Bed material entrained in the flue gas is typically removed by a cyclone in series with a fabric filter (FF) or an electrostatic precipitator (ESP).

Circulating fluidized-bed combustors are similar to bubbling-bed combustors except that the superficial velocities within the bed are 3 to 5 times higher than in bubbling-bed combustors. As a result, a physically well-defined bed is not formed; instead, solid particles are entrained with the transport air/combustion gases. Most of the solids are captured by a cyclone, and are continuously recirculated into the combustor. The solids still in the flue gas are captured by a downstream FF or an ESP.

Available information on the capital and operating costs of FBC's are insufficient to distinguish bubbling versus circulating bed designs.[6,7] Therefore, a single set of cost procedures has been developed.

2.1.4.2 Capital Cost Procedures. Table 2.1-4 presents the capital cost procedure for FBC's. All costs are estimated in December 1987 dollars. The procedures do not include the costs for FF's or ESP's. The costs for ESP's and FF's can be estimated using procedures discussed in Sections 2.2 and 2.3 of this report, respectively.

The direct and indirect cost equation shown in Table 2.1-4 is based on a vendor cost estimate for two combustors each rated at 450 tpd (i.e., total plant capacity of 900 tpd).[8] The vendor cost estimate included: combustor vessel, natural gas preheat system, forced draft and induced draft fans, tramp removal system, boiler, fuel and limestone storage and metering, multiclone, instrumentation and control including a boiler management system, ductwork, freight, and engineering and start-up supervision. However, the cost estimate provided by the vendor did not include cRDF processing and other equipment associated with the combustors such as front-end loaders, primary shredders, weight scales, magnetic separators, stacks, cooling tower, water treatment and supply, and steam turbine. Because the other equipment would be the same for both cRDF and FBC facilities, costs for the balance of plant in Table 2.1-3 (adjusted by size to 900 tpd using the 0.6 power cost rule) were added to the vendor cost estimates to estimate the total direct cost.[9] Indirect costs for general facilities and engineering fees were based on percentages of the direct cost estimated by Table 2.1-3.

To estimate the direct and indirect capital cost for other plant sizes, it is assumed that the economy of scale of FBC facilities is the same as for mass-burn and RDF facilities (0.39). Similar indirect installation costs would be incurred for these boiler types. A 20-percent process contingency is added to account for the relatively limited application of FBC to MWC's.[10]

2.1.4.3 Operating Cost Procedures. Table 2.1-5 presents the operating cost bases for FBC's. Table 2.1-6 presents operating cost equations derived from Table 2.1-5. The FBC operating procedures are divided into two process areas: combustor and the balance of plant, and cRDF processing. The direct operating cost bases for the combustor and balance of plant were based on information provided by one vendor for labor, electrical consumption, and water requirements. The maintenance materials costs were determined from cost data for coal-fired industrial FBC's.[11] Lime costs are based on the

TABLE 2.1-4. PROCEDURE FOR ESTIMATING CAPITAL COSTS FOR NEW FBC'S (December 1987 dollars)

Total Direct and Indirect Costs:[a]

$$\text{Costs, } 10^3\$ = 64{,}900 * \text{TPD} * (900/\text{TPD})^{0.39}$$

Process Contingency: 20% of total direct and indirect costs

Total Capital FBC Costs: Total direct and indirect costs + process contingency

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[a] TPD = plant municipal waste feed rate, tons/day.

TABLE 2.1-5. ANNUAL OPERATING COST PROCEDURES FOR NEW FBC'S

		References
Combustors and Balance of Plant (except coarse RDF processing area)		
Operating Labor:	10 man-years/year, 40 hours/week, and $12/hr for a 900 tpd plant	6, 12
Supervision:	3 man-years/year, 40 hours/week, and 30% premium over operating wage for a 900 tpd plant	6, 13
Maintenance Labor:	3 man-years/year, 40 hours/week, and 10% premium over operating wage for a 900 tpd plant	6, 14
Maintenance Materials:	3% of the total capital costs	11
Electricity:	3 MW power consumption for a 900-tpd plant and electricity costs of $0.046/kwh	6
Limestone:	$40/ton	15
Water:	3% blowdown rate calculated from steam production and $0.50/1,000 gal for water	6, 16
Waste Disposal:	$25/ton tipping fee and 99% combustible material in RDF and spent sorbent collected	6, 17
Overhead:	60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of the maintenance materials costs	18
Taxes, Insurance, and Administrative:	4% of the sum of the total capital costs	18
Capital Recovery of FBC Facility:	15 year life and 10% interest rate	19

Continued

TABLE 2.1-5 (CONCLUDED). ANNUAL OPERATING COST PROCEDURES FOR NEW FBC'S

		References
Coarse RDF Processing Area		
Total Operating and Maintenance:	4.4% of the total direct and indirect capital costs * ratio of operating to capital costs from Equation 6 in Section 2.1.2	1
Taxes, Insurance, and Administrative:	4% of the sum of the total capital costs	18
Capital Recovery of FBC Facility:	15 year life and 10% interest rate	19

TABLE 2.1-6. PROCEDURE FOR ESTIMATING ANNUAL OPERATING COSTS FOR FBC'S (December 1987 dollars)

Combustor and Balance of Plant (excludes coarse RDF processing area):

Operating labor (based on 10 man-years, 40 hours/week, $12/hr):

$OL = 10 * 40 * 52 * 12 * (TPD/900) = 277.3 * TPD$

Supervision (based on 3 man-years/year, 40 hours/week, 30% wage rate premium over the operating labor wage):

$SPRV = 3 * 40 * 52 * 12 * 1.3 * (TPD/900) = 108.2 * TPD$

Maintenance labor (based on 3 man-years/year, 40 hours/week, 10% wage rate premium over the operating labor wage):

$ML = 3 * 40 * 52 * 12 * 1.1 * (TPD/900) = 91.5 * TPD$

Maintenance materials: 3% of the total capital costs

Electricity (based on 3 MW power consumption, and electricity rate of $0.046/kwh):

$ELEC = 0.153 * TPD * HRS$

Limestone (based on $40/ton for limestone):

$LIMESTONE = 0.02 * LFEED * HRS * N$

Water (based on 3% blowdown rate and $0.05/1,000 gal):

$WC = 1.86 \times 10^{-6} * STM * HRS$

Waste disposal (based on tipping fee of $25/hr):

$AD = 1.25 \times 10^{-2} * N * HRS * WDR$

Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus 60% of maintenance materials costs

Taxes, Insurance, and Administrative Charges:

4% of the total capital cost

Capital Recovery (based on 15 year life and 10% interest rate):

13.15% of the total capital cost

Continued

TABLE 2.1-6 (CONCLUDED). PROCEDURE FOR ESTIMATING ANNUAL OPERATING COSTS FOR FBC'S (December 1987 dollars)

Coarse RDF Processing Area:

Total Operating and Maintenance Costs (TOT O&M):

$\text{TOT O\&M} = 4.4 \times 10^{-4} * (12.5 - 0.00115 * \text{TPD}) * \text{TDI}$

Taxes, Insurance, and Administrative Charges:

4% of the total capital cost

Capital Recovery (based on 15 year life and 10% interest rate):

13.15% of the total capital cost

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[a] OL = operating labor costs, $/yr
SPRV = supervision costs, $/yr
ML = maintenance labor costs, $/yr
ELEC = electricity costs, $/yr
HRS = hours of operation per year
LIMESTONE = limestone costs, $/yr
LFEED = limestone feed rate per unit, lb/hr
N = number of combustors
WC = water costs, $/yr
STM = plant steam production, lb/hr
AD = waste disposal costs, $/yr
WDR = waste disposal rate per unit (bottom and fly ash collected), lb/hr
TPD = plant municipal waste feed rate, tons/day
TDI = total direct and indirect capital costs for FBC plant, $

amount of limestone injected at a cost of $40/ton. This cost is based on a limestone freight-on-board cost of $20/ton and a transportation cost of $20/ton, assuming a handling rate of $0.04/ton-mile hauling distance and a 500-mile hauling distance. The cost for waste disposal is determined from the amount of solids removed by the FBC plus additional fly ash collected by the downstream particulate control device.

All cost rates are based on December 1987 dollars. The operating labor wage is the average from those wages in the Department of Commerce Survey of Current Business for private nonagricultural payrolls and EPRI's Technical Assessment Guide.[20,21] Electricity rates are from the Energy Information Administration, Monthly Energy Review.[22]

The annual operating cost procedures for RDF processing facilities in Section 2.1.3.3 were used to estimate the annual operating costs for the cRDF processing area. Operating costs for the cRDF processing area are estimated to be 4.4 percent of the product of total FBC capital cost and the ratio of operating-to-capital costs for RDF facilities from Equation 6 in Section 2.1.2.3.

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2.2 ELECTROSTATIC PRECIPITATORS

2.2.1 Overview of Technology

Electrostatic precipitators are used to control PM emissions from MWC's. In this process, flue gas flows between a series of high voltage discharge electrodes and grounded metal plates. Negatively charged ions formed by this high voltage field attach to particulate in the flue gas, causing the charged particles to migrate toward the grounded plates. Charged particles that collect on the grounded plates are periodically removed by rapping or washing. Key ESP design and operating characteristics influencing ESP performance are particulate size and resistivity, specific collection area (SCA, equal to the total surface area of the collection plates divided by the flue gas flow rate), and the number of ESP fields. When the plates are cleaned, some of the collected particulate is reentrained in the flue gas. To ensure good particulate collection efficiency during plate cleaning and electrical upsets, ESP's have several fields located in series along the direction of flue gas flow that can be energized and cleaned independently. Particles reentrained when the dust layer is removed from one field can be recollected in a downstream field.

2.2.2 Capital Cost Procedures

2.2.2.1 Direct costs. Information on direct equipment costs is available for ESP's at three PM control levels (0.01, 0.02, and 0.03 gr/dscf at 12 percent CO_2) for mass-burn, modular, and RDF facilities ranging in size from 100 to 3,000 tpd total plant capacity.[1] These cost estimates, based on data provided by eight manufacturers, are presented in Table 2.2-1.

The equipment cost data were correlated with total plate area. Figure 2.2-1 illustrates the "best fit" equation for the data from all of the ESP vendors except for one. The data from manufacturer "D" in Table 2.2-1 at a flue gas flowrate of 245,230 acfm were not included in the analysis because the cost data varied significantly from the rest of the data (refer to the three solid data points in Figure 2.2-1). Data from manufacturers F and G in Table 2.2-1 were excluded in Figure 2.2-1, because flue gas flowrates for the large mass-burn units were not reported. The resultant "best fit" equation using this approach is:

TABLE 2.2-1. VENDOR QUOTES FOR ESP EQUIPMENT COSTS (IN 1000$ AUGUST 1986)[a]

Vendor	Furnace type[a]	Flue gas flowrates, acfm	Outlet PM emissions, gr/dscf		
			0.03	0.02	0.01
A	MOD	54,105	240	250	310
A	MOD	86,568	280	290	390
B	MOD	86,568	325	325	325
C	MB	24,523	253	423	440
D	MB	24,523	410	450	570
E	MB	76,000	NA[b]	640	NA
C	MB	NA	503	828	980
D	MB	NA	1,470	1,640	2,310
F	MB	245,230	1,313	NA	1,750
G	MB	240,000	NA	1,813	2,188
B	MB	190,031	475	475	545
H	MB	126,687	567	576	617
H	RDF	130,843	580	645	781
A	RDF	130,843	780	880	890
H	RDF	196,264	768	832	977
A	RDF	196,264	910	970	980

[a] MOD = modular; MB= mass-burn, and RDF = refuse-derived fuel.

[b] NA = not available.

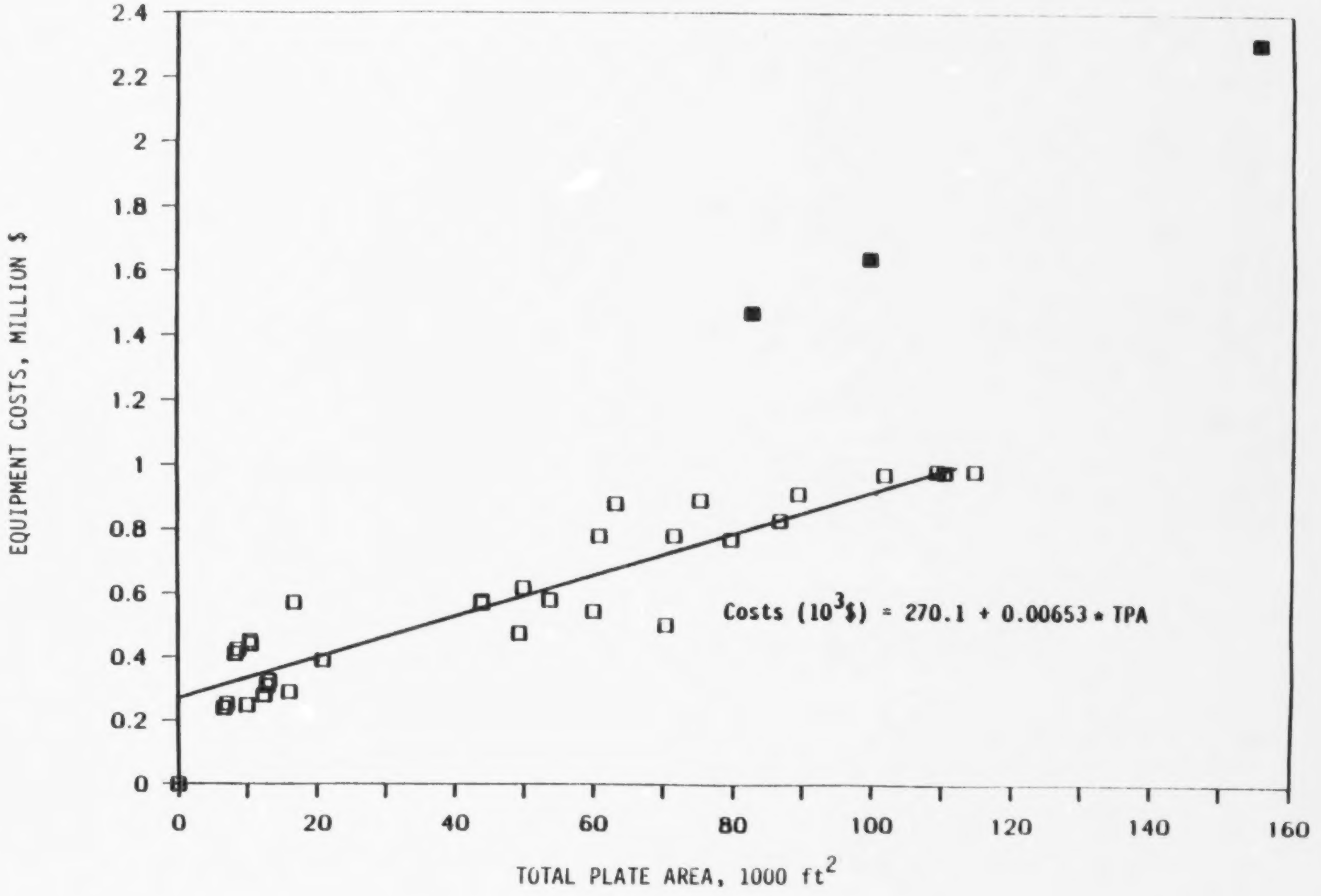

Figure 2.2-1. Correlation of ESP equipment costs (in August 1986 dollars) from ESP manufacturers and total plate area.

$$\text{Purchase equipment cost, } 10^3\ \$ = 305.2 + 0.00738 * \text{TPA, in December 1987 dollars} \quad (1)$$

where: TPA is the total plate area in ft^2 calculated as the product of the SCA in ft^2/1000 acfm and the flue gas flowrate in 1000 acfm.

The flue gas flowrate is assumed to be 125 percent of the design flowrate to accommodate variations in feed waste composition and operating conditions.[2] Equation 1 was derived using the Chemical Engineering Plant Index to update the cost equation shown in Figure 2.2-1 to December 1987 dollars and including the cost for taxes and freight. Taxes and freight were estimated at 8 percent of the equipment cost.[3]

To estimate the required SCA for new units, the following approach was taken. Data on SCA's were provided by eight manufacturers, as shown in Table 2.2-2.[4] From this table, the average SCA was calculated for each PM removal efficiency, as shown in Table 2.2-3. The average SCA's from Table 2.2-3 were correlated with PM collection efficiency using the Deutsch-Anderson equation. The Deutsch-Anderson equation is frequently used to predict ESP performance.[5] Using the form of the Deutsch-Anderson equation, the following equation is derived:

$$\text{PM collection efficiency, \%} = 100 - 101.89 * \exp(-0.0112 * \text{SCA}) \text{ or}$$

$$\text{SCA} = -(89.29) * \ln [(100 - \text{PM collection efficiency, \%})/101.89] \quad (2)$$

where SCA = specific collection area, ft^2/1000 acfm

Figure 2.2-2 presents the "best fit" equation for the average SCA data.

Both equations (Equations 1 and 2) reasonably fit the data. The coefficients of determination (R^2) were 0.86 for Equation 1 and 0.97 for Equation 2. Part of the scatter not explained by Equation 1 may be due to differences in equipment included in different vendor estimates.

Equations 1 and 2 apply to field-erected ESP's with total plate areas above 6,500 ft^2 and flue gas flowrates above 30,000 acfm. ESP's applied to

TABLE 2.2-2. SPECIFIC COLLECTION AREA (SCA) REPORTED BY THE ESP MANUFACTURERS

Manufacturer	Flue Gas Flowrate, acfm	Inlet PM Loading, gr/dscf	SCA, ft^2/1000 acfm, for outlet loading of 0.03	0.02	0.01
A	54,015	0.11	121	182.8	235.3
A	86,568	0.11	142	184.4	239.3
B	86,568	0.11	150	150	150
C	24,523	1.72	285.5	346.7	423.7
D	24,523	1.72	332.1	419.6	675.9
E	76,000	NA	NA	NA	NA
C	245,230	1.72	286	352.2	444.9
D	245,230	1.72	335.7	405.6	634
F	NA[a]	1.72	325	375	450
G	NA	1.72	360	400	500
B	190,031	1.72	258	258	314
I	NA	1.72	400	500	NA
H	126,687	1.72	345	NA	392
H	130,843	4.63	408.6	NA	544.8
A	130,843	4.63	462.6[b]	480.8	573[b]
H	196,264	4.63	404.5	NA	561.3
A	192,264	4.63	463[b]	527.9	595.1[b]

[a] NA = not available.

[b] These values were not used to estimate averages SCA values reported in Table 2.2-3.

TABLE 2.2-3. AVERAGE SPECIFIC COLLECTION AREA (SCA) CALCULATED FROM THE MANUFACTURERS' DATA

Inlet PM Loading, gr/dscf at 12% CO_2	Outlet PM Loading, gr/dscf at 12% CO_2	PM Removal Efficiency, Precent	SCA, ft^2/1000 acfm
0.11	0.03	72.7	138 (3)[a]
0.11	0.02	81.8	172 (3)
0.11	0.01	90.9	208 (3)
1.72	0.03	98.3	332 (9)
1.72	0.02	98.8	397 (8)
1.72	0.01	99.4	500 (8)
4.63	0.03	99.4	406 (2)
4.63	0.02	99.6	504 (2)
4.63	0.01	99.8	553 (2)

[a] Number in parantheses indicates the number of data points used for the average.

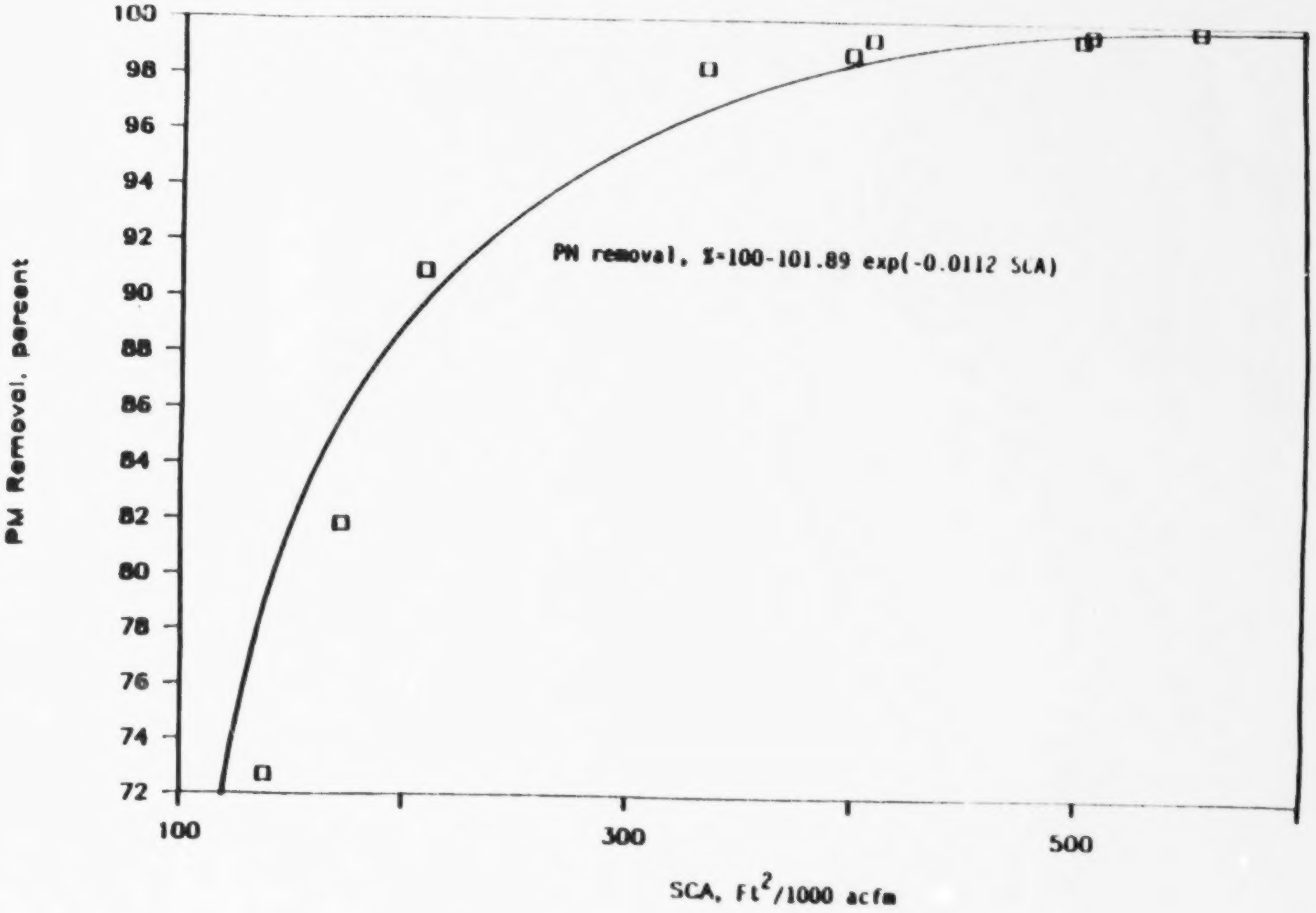

Figure 2.2-2. Relationship between ESP manufacturers' specific collection area and particulate matter removal.

smaller modular combustor facilities generally are shop-assembled and are installed at the facility at a minimal cost. To estimate the costs of this type of ESP, cost data from one manufacturer were analyzed using the same approach as for field-erected ESP's.[6] The "best fit" equation relating purchase equipment costs to total plate area (TPA) is:

$$\text{Purchase equipment costs, } 10^3\$ = 96.3 + 0.015 * \text{TPA, in December 1987 dollars} \qquad (3)$$

$$R^2 = 0.86$$

Figure 2.2-3 presents the "best fit" equation for the data from this manufacturer. Costs received from another ESP manufacturer after Equation 3 was developed are similar to those used to develop Equation 3.[7]

To estimate the required SCA, the SCA calculated from the manufacturer cost data was correlated with PM collection efficiency using the Deutsch-Anderson equation:

$$\text{PM collection efficiency, \%} = 100 - 79.6 * \exp(-0.0035 * \text{SCA}) \text{ or}$$

$$\text{SCA} = -(285.7) * \ln[(100-\text{PM collection efficiency, \%})/79.6] \qquad (4)$$

$$R^2 = 0.90$$

Figure 2.2-4 presents the "best fit" equation for the SCA data.

Table 2.2-4 summarizes the procedure for estimating total capital cost for ESP's using the above four equations. The SCA required to achieve a given PM collection efficiency is estimated using either equation 2 or 4. Purchased equipment costs for the ESP can then be obtained using either Equation 1 or 3. For the costs of additional ESP units, the costs of a single ESP are multiplied by the number of required units. Procedures for estimating the costs of ductwork and fan, and installation direct costs are also presented in Table 2.4-4.[8]

2.2.2.2 Indirect and Other Costs. The cost factors for estimating indirect costs for field-erected ESP's are based on those presented in established EPA cost procedures.[9,10] Because installation and engineering

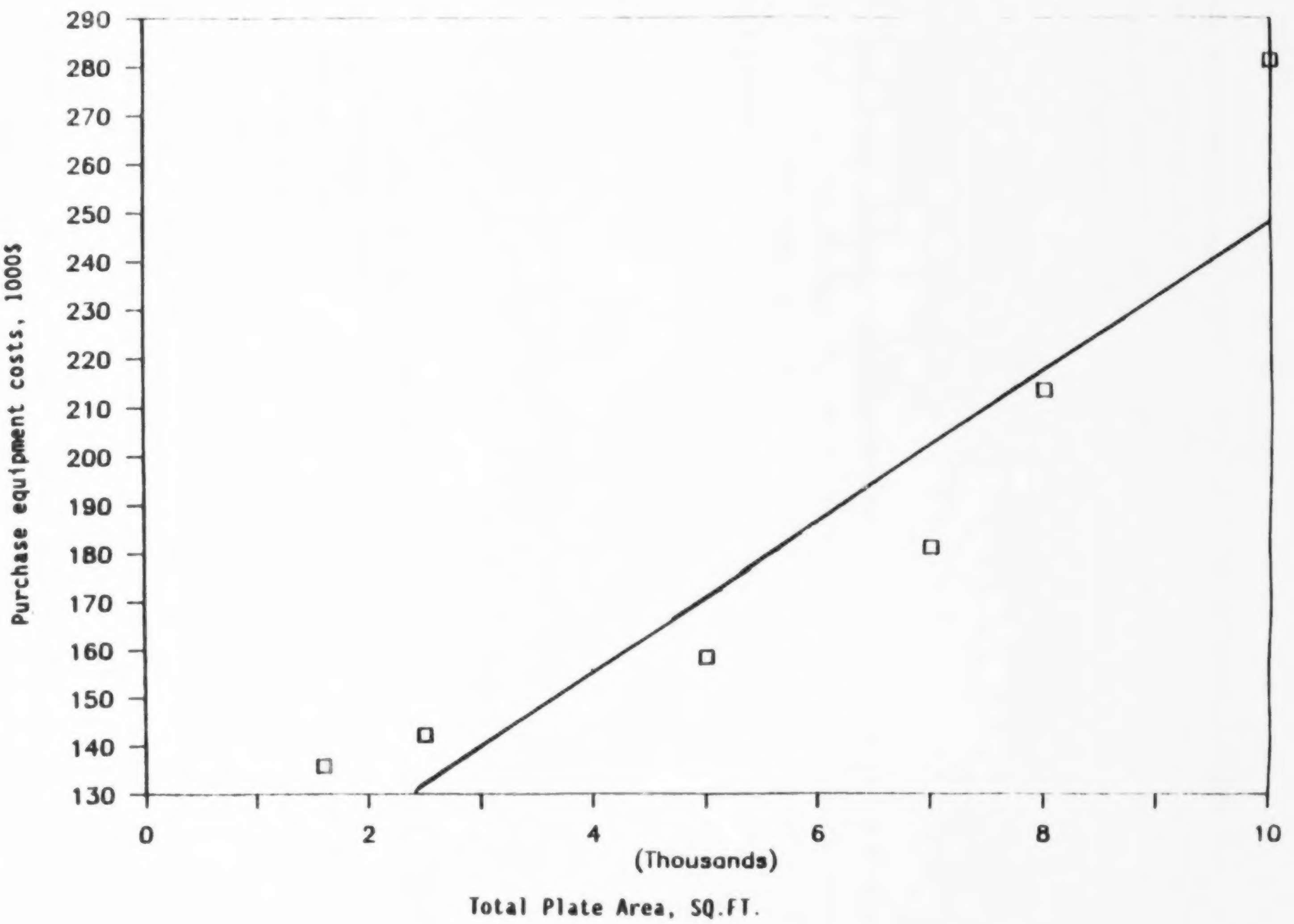

Figure 2.2-3. Correlation of ESP purchase equipment costs with total plate area for modular ESP's.

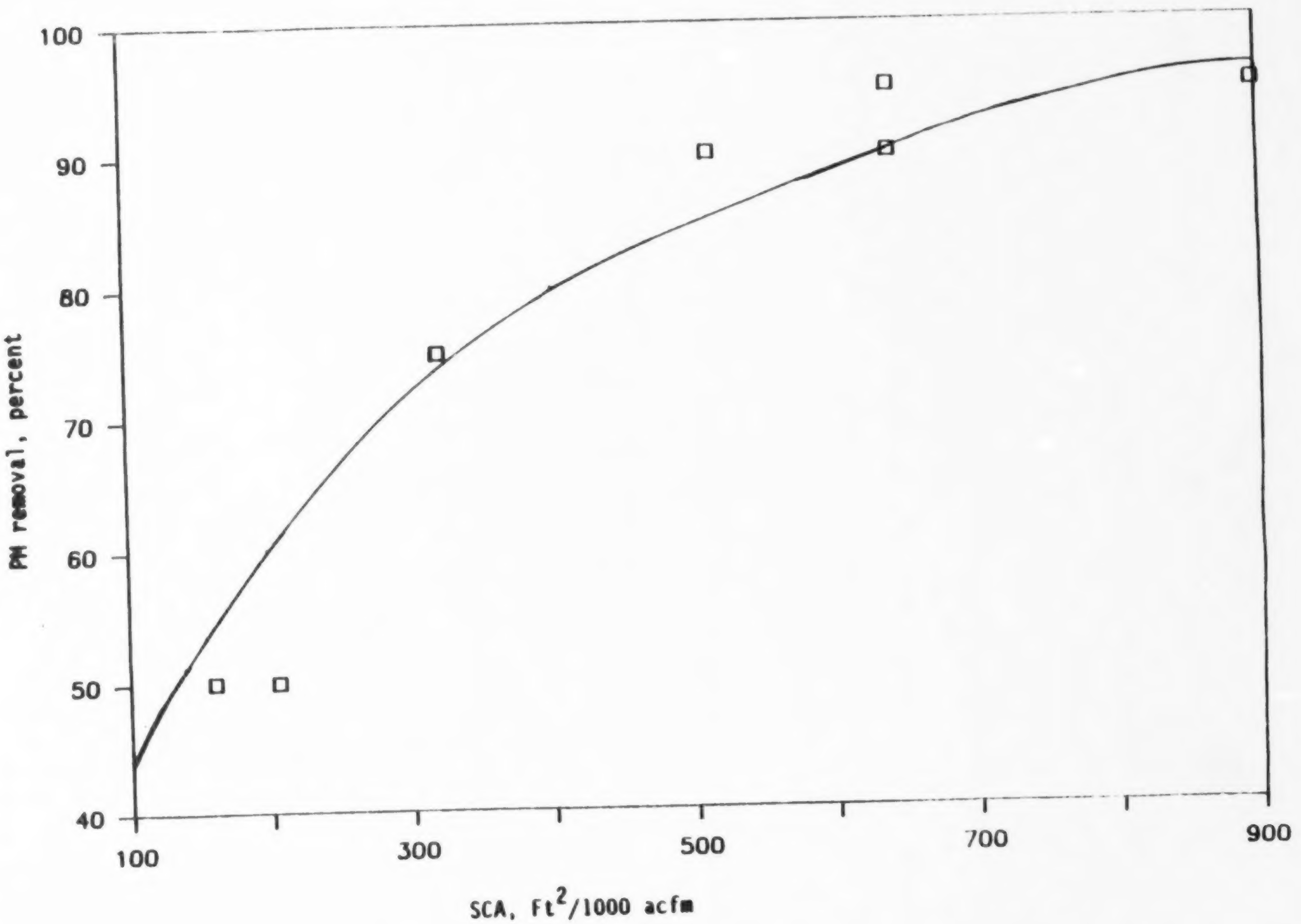

Figure 2.2-4. Relationship between specific collection area and particulate matter removal for modular ESP's.

TABLE 2.2-4. COST PROCEDURES FOR ESTIMATING CAPITAL COSTS FOR ESP'S ON NEW PLANTS[a]

Purchased Equipment Costs (December 1987 dollars):

Single Field-erected ESP unit[b]: Costs, $10^3\$ = 305.2 + 0.00738 * TPA$
$TPA = SCA * Q/1000$
PM efficiency $\% = 100-101.89 \exp(-0.0112 * SCA)$

Single Shop-assembled ESP unit[b,c]: Costs, $10^3 = 96.3 + 0.015 * TPA$
PM efficiency $\% = 100-79.6 \exp(-0.0035 * SCA)$

Multiple Units: Costs = N * Costs for single ESP unit

Ductwork: $\text{Costs} = 0.7964 * L * Q^{0.5}$

Fan: $\text{Costs} = 1.077 * Q^{0.96}$

Installation Direct Costs = 67% of equipment costs

Indirect Costs = 54% of Purchase Equipment Costs for field-erected ESP's
= $14,000 for shop-assembled ESP's

Contingency = 3% of the Purchase Equipment Costs

Total Capital Investment = Purchased Equipment Costs + Installation Direct Costs + Indirect Costs + Contingency costs are less for a shop-assembled ESP than for a field-erected ESP, indirect costs for shop-assembled ESP's are based on the manufacturer's estimate of $14,000.[11] Costs reported in August 1986 dollars were updated to December 1987 dollars using the Chemical Engineering Plant Cost Index for all equipment.

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[a] TPA = total plate area, ft^2
SCA = specific collection area, ft^2/1000 acfm
Q = 125 percent of the calculated flue gas flowrate, acfm
L = Duct length, ft
N = Number of ESP units

[b] Includes taxes and freight of 8 percent of the ESP equipment costs.

[c] Applies only to modular combustors whose flue gas flowrate Q is less than 30,000 acfm.

2.2.3 Operating Cost Procedures

Table 2.2-5 presents the procedure for estimating annual operating costs for ESP's. The procedures and factors shown for estimating the various components of annual operating costs and the references for each are listed in Table 2.2-5. Operating costs are presented in December 1987 dollars.

To the extent that data are available, cost rates are based on actual rates in December 1987 dollars. Operating labor wage rate is the average from those obtained from the U. S. Department of Commerce, in its Survey of Current Business for private non-agricultural payrolls and the EPRI's Technical Assessment Guide.[12,13] Electricity rates were obtained from the Energy Information Administration, in its Monthly Energy Review.[14] An ash disposal cost rate of $25/ton was used, since typical ash disposal rates (tipping fees) are between $20 and $30/ton.[15]

TABLE 2.2-5. COST PROCEDURES USED TO ESTIMATE ANNUAL OPERATING COSTS FOR ESP'S ON NEW UNITS

		References
Operating Labor:	1 man-hour/shift[a]	16
Supervision:	15% of operator labor costs	17
Maintenance:		
Labor	0.5 man-hour/shift, 10% wage rate premium over operating labor wage	17, 18
Materials	1% of the total capital costs	17
Electricity:	1.5 watts/ft^2 collection area	19
	0.5 inch pressure drop W.C.	20
Ash Disposal:	$25/ton	21
Overhead:	60% of the sum of all labor costs (operating, supervisory, and maintenance) and 60% of the maintenance material.	22
Taxes, Insurance, and Administrative Charges:	4% of total capital costs	22
Capital Recovery:	15 year life and 10% interest rate	15

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[a] Labor requirement in the range reported by Reference 16 (from 1/2 to 2 man-hour/shift).

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19. Reference 17, p. 3-18.

20. Reference 17, p. 5-2.

21. Reference 10, p. 2-29.

22. Reference 10, p. 2-31.

2.3 DRY SORBENT INJECTION

2.3.1 Overview of Technology

Dry sorbent injection is being examined as a control option for achieving moderate acid gas control and indirectly increasing dioxin control for MWC's. Two basic variations of this technology exist:

- furnace sorbent injection in which alkali sorbent is injected through the overfire air ports into the furnace, and
- duct sorbent injection in which the sorbent is injected into either a duct or a reactor vessel upstream of the particulate control system.

Particulate control following sorbent injection can be accomplished by either an ESP or fabric filter.

Sorbent injection technologies have been used commercially on MWC's in Europe and Japan since 1979.[1] Japanese duct injection technology generally uses a high-temperature (approximately 500°F) ESP for particulate matter collection. European duct injection technology incorporates a fabric filter (FF) for particulate control with typical FF inlet temperatures of 350°F. Furnace and duct sorbent injection systems have recently been installed and tested at several MWC's in the U. S. In addition, significant testing of furnace and duct injection applied to coal-fired systems for SO_2 control has occurred. However, data on the comparative performance and the cost of different sorbent injection approaches for MWC's are limited.

The basic chemistry for acid gas control is the reaction of calcium or sodium sorbent with HCl and SO_2 to form chloride, sulfite, and sulfate salts. The degree of acid gas control is a function of sorbent feed rate, the extent of flue gas and sorbent mixing, the flue gas temperature, and the PM control device. For moderate levels of acid gas control, sorbent can be injected directly into the furnace or the flue gas duct. For higher levels of control, a separate reactor vessel can be used that is designed to enhance flue gas and sorbent mixing and provide additional reaction time. Flue gas humidification with water sprays or additional heat recovery in an economizer or air preheater can be used to reduce flue gas temperature. Procedures for estimating the costs of flue gas temperature reduction using humidification are presented in Section 3.5.

Based on similarities in equipment requirements, the capital costs for furnace and duct injection are expected to be generally similar. As a result, a "generic" cost procedure was developed to estimate the capital and operating costs for both types of sorbent injection. Major equipment associated with both technologies consists of a storage silo, a pneumatic feeding system for transferring sorbent from the storage silo to feed bins, feed bins with gravimetric metering systems, and pneumatic sorbent injectors.[2] For duct sorbent injection, a venturi or a reaction vessel with mixing baffles is provided to ensure adequate gas-to-sorbent mixing. For furnace sorbent injection, sorbent will be injected through the overfire air ports or separate injection ports in the combustor. For new systems, a FF is assumed for PM control because of its enhanced acid gas and dioxin removal capabilities compared to an ESP.

Primary operating costs include labor, maintenance materials, electricity, and sorbent. Labor, maintenance material, and electricity cost are expected to be generally similar for both duct and furnace sorbent injection. Because of the greater amount of data on calcium-based sorbents, the cost procedures assume use of hydrated lime ($Ca(OH)_2$). For furnace injection, limestone ($CaCO_3$) or lime (CaO) can be used which may be less expensive.

2.3.2 Capital Cost Procedures

The direct capital cost of sorbent injection equipment depends on the flue gas and sorbent flowrates. These two parameters, in turn, depend on MSW feed rate and composition, excess air levels, flue gas temperature, sorbent quality and utilization rate, and emission control requirements. Based on a simple material balance that assumes all of the sulfur and chlorine in the MSW are converted to SO_2 and HCl, sorbent throughput requirements can be calculated using the following equation:

$$\underset{\text{(lb/hr)}}{\text{Lime Feed Rate}} = 74.1 * (2{,}000/24) * \text{TPD} * [\%\text{S}/32 + \%\text{Cl}/71] * \text{CAG/PURITY}$$

where:

TPD = tons per day of MSW, based on 125 percent of the design capacity to accommodate variations in feed waste composition and operating conditions[3],

%S = percent sulfur in the MSW,

%Cl = percent chlorine in the MSW,

CAG = calcium-to-acid gas molar ratio (i.e., stoichiometric ratio), and

PURITY = weight percent of calcium in the lime.

Based on available data for duct sorbent injection, a value of 2 for CAG is expected to achieve removal efficiencies of 80 percent for HCl and 40 percent for SO_2.[4] For furnace sorbent injection, a CAG value of 2 is expected to achieve 70 percent removal of HCl and 70 percent removal of SO_2.[5] PURITY is assumed to be 90 percent.

Procedures for calculating direct capital costs for the individual major equipment items shown in Table 2.3-1 were derived from data in standard cost estimating manuals and manufacturer estimates.[6,7,8] Cost for a reactor vessel is based on a vaned, stainless steel tank with one second of flue gas residence time.[9] Installation costs are assumed to be 30 percent of the equipment costs.[10] Indirect costs, also shown in Table 2.3-1, are calculated as a percentage of total direct costs. These indirect cost rates are the same as those used for estimating the indirect capital costs of a spray dryer/FF (presented in Section 2.4-1). The equations in Table 2.3-1 estimate costs in December 1987 dollars. The costs were escalated to December 1987 dollars using the Chemical Engineering Plant Cost Index for all equipment.

Capital costs for pulse jet FF's with a net air-to-cloth ratio of 4:1 are estimated using equations for single units.[11] Direct and indirect capital costs as a function of flue gas flowrate can be estimated from these equations. Installation and indirect costs for FF's are 72 and 42 percent of the equipment cost, respectively.[12] The flue gas flowrate is based on 125 percent of the design flowrate to accommodate variations in feed waste composition and operating conditions. The costs for FF and auxiliary equipment are in December 1987 dollars. Contingency is included to account for unforeseen costs (50 percent of the direct and indirect cost) during installation and start-up due to the relative lack of operating experience of dry sorbent injection systems applied to MWC's.[13] Projected equipment life is 15 years.[14]

TABLE 2.3-1. PROCEDURES FOR ESTIMATING CAPITAL COSTS FOR DRY SORBENT INJECTION[a,b]

Purchased Equipment Costs, 10^3 \$

1. Lime Storage Silo with Vibrator, Baghouse, and Flow Control Value (Based on 15-day lime supply)

   - o For storage volumes (V) less than or equal to 2,300 ft^3 (one storage silo per plant),
     Costs = 1.05 * (34.2 + 0.016V) * RF

   - o For storage volumes between 2,300 and 4,600 ft^3 (two storage silos per plant),
     Costs = 2.10 * (34.2 + 0.016V) * RF

   - o For storage volume greater than 4,600 ft^3 (two storage silos per plant),
     Costs = 2.10 * (63 + 0.0038V) * RF

2. Feed Bins

   - o For duct sorbent injection (one feed bin per combustor),
     Costs = 0.0906 * N * RF * $(SF)^{0.6145}$

   - o For furnace sorbent injection (two feed bins per combustor),
     Costs = 0.1812 * N * RF * $(SF)^{0.6145}$

3. Gravimetric Feeders

   - o For duct sorbent injection (one feeder per combustor),
     Costs = 1.024 * (0.000289 SF + 9.293) * N * RF

   - o For furnace sorbent injection (two feeders per combustor),
     Costs = 2.048 * (0.000289 SF + 9.293) * N * RF

4. Pneumatic Conveyor (Based on 400 Feet Length)

   Costs = 1.05 * (26.4 + 0.0073 SF + 0.4 * [12.8 + 11.23 $SF^{0.23}$]) N * RF

5. Injection Ports

   - o For duct sorbent injection (one injection port per combustor),
     Costs = 1.05 * (22.2 + 0.0014 SF) * N * RF

   - o For furnace sorbent injection (two injection ports per combustor),
     Costs = 2.10 * (22.2 + 0.0014 SF) * N * RF

(continued)

TABLE 2.3-1. (Continued)

6. Reactor Vessel (optional for duct sorbent injection to increase flue gas and sorbent contact): $\text{Costs} = 34 * (Q * 1.25/6{,}150)^{0.64}$

7. Fabric Filter[c]

   $\text{Costs} = 0.1482 * N * RF * Q^{0.7043}$

8. Induced Draft Fan

   $\text{Costs} = (1.167 * N * RF * Q^{0.96})/1{,}000$

9. Ductwork

   $\text{Costs} = (0.8627 * N * RF * L * Q^{0.5})/1{,}000$

Installation Direct Costs

= 30% of dry sorbent injection equipment costs + 72% of fabric filter and auxiliary equipment costs

Indirect Costs

= 33% of direct costs (equipment + installation costs) for dry sorbent injection + 42% of equipment cost for the fabric filter and auxiliary equipment

Contingency

= 50% of the sum of direct and indirect costs

Total Capital Costs

= Total Direct Costs + Indirect Costs + Contingency Costs

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[a] All costs are estimated in December 1987 dollars.

[b] SF = lime feed rate per unit, lb/hr
Q = 125 percent of the actual flue gas flowrate to the fabric filter per unit, acfm
V = lime storage silo volume for the plant, ft^3
N = number of units
RF = Retrofit factor. For retrofit applications, use retrofit factor of 1.1 for sorbent injection equipment. Retrofit factors for fabric filter and auxiliary equipment are obtained from Table A-14. For new units, RF = 1.0.
L = duct length, feet

[c] Fabric filters are used for new applications and for retrofit applications where no ESP exists. For plants with existing ESP's, costs for upgrading the ESP are estimated from Table B-4.

2.3.3 Operating Cost Procedures

Table 2.3-2 presents procedures for estimating operating costs for dry sorbent injection alone. Operating costs for humidification are presented in Section 3.5. The operating and maintenance labor requirements and maintenance materials for sorbent injection are based on typical values for coal-fired boilers. Electricity costs are based on electrical requirements to operate the pneumatic feed systems. Lime costs are based on the amount of lime injected. Equations for electricity and lime were taken from Reference 10.

Table 2.3-3 presents procedures for estimating operating costs for FF's. The operating and maintenance labor requirements are based on those from established EPA procedures, with the exception of maintenance materials. Because maintenance material requirements for FF's can vary directly with the size of the unit, maintenance material costs are assumed to be calculated at five percent of the direct capital costs. This percentage is the same one used for dry sorbent injection to estimate maintenance material costs. The cost of bag replacement assumes a 2-year bag life, which is typical for FF's. A gross air-to-cloth ratio of 3:1 is used.

Electricity to operate the I.D. fan is calculated using a total pressure drop of 12.5 inches of water, 7 inches of water across the FF and 5.5 inches for the additional ductwork and dry sorbent injection. The cost of compressed air for the pulse jet FF's is estimated from established EPA procedures. The costs for solids disposal are determined from the amount of solids collected by the FF and a tipping fee of $25/ton.

All cost rates are based on December 1987 dollars. The operating labor wage is the average from those obtained from the Department of Commerce Survey of Current Business for private nonagricultural payrolls and EPRI's Technical Assessment Guide.[15,16] Electricity rates will be obtained from the Energy Information Administration, Monthly Energy Review.[17] Operating hours per year can be varied to meet model plant specifications.

Indirect operating costs such as taxes, insurance, and administrative charges are based on percentages of the capital costs. Payroll and plant overhead are based on a percentage of the labor and material costs.

TABLE 2.3-2. ANNUAL OPERATING COST PROCEDURES FOR DRY SORBENT INJECTION FOR NEW MWC's[11]

		References
Operating Labor:	2 manhour/shift	14
Supervision:	15% of operator labor costs	18
Maintenance Labor:	0.5 manhour/shift, 10% premium over operating labor wage	18, 19
Materials:	5% of total direct costs	20, 21
Electricity:	(52.56 * (lime feed rate[a]) + 251,850) * (electricity costs) * (hours of operation/8,760)	22
Lime:	4.38 * (lime feed rate[a]) * (lime cost) * (hours of operation/8,760)	22
Overhead:	60% of the sum of all labor costs (operating, supervisory, and maintenance) plus maintenance material	23
Taxes, Insurance, and Administrative Charges:	4% of total capital costs	23

[a] Lime feed rate in lb/hr is based on 100 percent capacity of waste processed.

TABLE 2.3-3. ANNUAL OPERATING COST PROCEDURES FOR FABRIC FILTERS FOR NEW MWC's

		References
Operating Labor:	2 manhour/shift	24
Supervision:	15% of operator labor costs	24
Maintenance Labor:	1 manhour/shift, 10% wage rate premium over operating labor wage	18, 24
Materials:	5% of direct capital costs	20
Bag Replacement:	\$1.35/ft^2 for teflon coated fiberglass; 2-year life	25
Electricity:	Calculated based on fan requirements for inches of water pressure drop across FF	26
Compressed Air:	2 scfm/1,000 acfm flue gas	27
Solid Waste:	Apply appropriate tipping fee in \$/ton (Assume \$25/ton)	28
Overhead:	60% of the sum of all labor costs (operating, supervisory, and maintenance) plus materials	23
Taxes, Insurance, and Administrative Charges:	4% of total capital costs	23
Capital Recovery:	15-year life and 10% interest rate	29

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2.4 SPRAY DRYING WITH EFFICIENT PARTICULATE CONTROL

2.4.1 Overview of Technology

Spray drying is designed to control SO_2 and HCl emissions. When used in combination with an efficient particulate control system, spray drying can also control CDD/CDF, PM, and metals emissions. In the spray drying process, lime slurry is injected into a spray dryer (SD) vessel. The water in the slurry evaporates to cool the flue gas, and the lime reacts with acid gases to form salts that can be removed by a PM control device. The simultaneous evaporation and reaction increases the moisture and particulate content in the flue gas. The particulate exiting the SD vessel contains fly ash plus calcium salts, water, and unreacted lime.

Spray drying is commonly used in combination with either a fabric filter (FF) or an electrostatic precipitator (ESP) for PM control. Both combinations have been used for MWC's in the United States, although SD/FF systems are more common and may be more effective for CDD/CDF, PM, and metals control. Two basic designs of FF's are available: reverse air and pulse jet. In a reverse air FF, flue gas flows through unsupported filter bags, leaving the particulate on the inside of the bags. The particulate builds up to form a particulate filter cake. Once an excessive pressure drop across the filter cake is reached, air is blown through the filter in the opposite direction, the filter bag collapses, and the filter cake falls off and is collected. In a pulse jet FF, flue gas flows through supported filter bags leaving particulate on the outside of the bags. To remove the built-up particulate filter cake, compressed air is introduced through the inside of the filter bag, the filter bag expands and the filter cake falls off and is collected. The cost procedures are based on pulse jet FF systems.

2.4.2 Capital Cost Procedures

Vendor capital cost estimates for SD systems combined with either an ESP or a FF applied to three types of MWC's (mass-burn, modular, and RDF) were obtained for systems designed to achieve 90 percent HCl and 70 percent SO_2 removal and PM emissions of 0.01, 0.02, and 0.03 gr/dscf at 12 percent CO_2.[1]

A cost comparison of SD/FF and SD/ESP systems designed to achieve a PM emission rate of 0.01 gr/dscf at 12 percent CO_2 is presented in Appendix A for two mass-burn facility capacity sizes (250 and 3,000 tons/day of MSW). This comparison indicates that, at this PM control level, costs for SD/FF and SD/ESP systems are very similar, with the annualized costs for SD/FF's being slightly lower than for SD/ESP's. Although cost procedures presented in this section focus on SD/FF systems, they are representative of costs for SD/ESP systems.

Cost procedures for stand-alone SD systems (i.e., without a FF) are presented in Section 3.6. These procedures were developed based on the SD/FF data plus supplemental cost quotes from three SD manufacturers. These cost procedures are intended to assist in evaluating methods to retrofit SD systems at existing plants already equipped with efficient PM control devices.

2.4.2.1 Direct Costs. Direct costs for an SD/FF system include purchased equipment cost for an SD, FF, induced draft (I.D.) fan, and ducting. The SD components include a reaction vessel, atomizer, lime feed preparation equipment, and solids handling equipment. The SD is sized based on a stoichiometric ratio (moles of calcium per mole of both SO_2 and HCl in the flue gas entering the spray dryer) of 1.5:1. The FF cost is based on a pulse-jet type unit operated at a net air-to-cloth ratio of 4:1 and a gross air-to-cloth ratio of 3:1.

Costs for single SD/FF units were based on cost data provided by two manufacturers as shown in Table 2.4-1. The data from these two manufacturers were used to estimate installed capital costs of SD/FF systems for all furnace types and are plotted as a function of flue gas flowrate in Figure 2.4-1.[2] The costs are approximately the same for any combustor type at the same flowrate. There are two reasons for this. First, the cost of the FF is assumed to be sensitive only to flue gas flowrate and is unaffected by PM grain loading. Second, the inlet SO_2 and HCl concentrations in the flue gas were assumed to be essentially the same for all facility types. Inlet SO_2 and HCl concentrations primarily depend on the MSW composition (particularly sulfur and chlorine contents) and MSW heating value. The values for these three factors assumed for the three facility types result in approximately the same SO_2 and HCl concentrations.

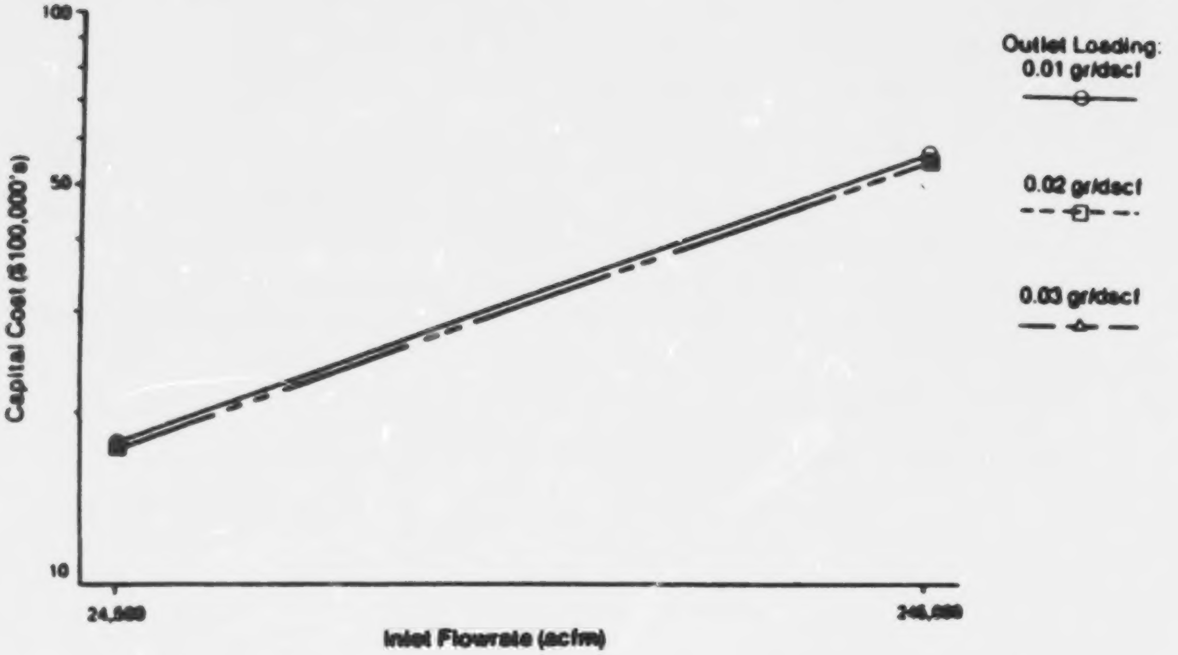
TABLE 2.4-1. VENDOR QUOTES FOR SPRAY DRYER/FABRIC FILTER TOTAL CAPITAL COSTS (IN $1,000 AUGUST 1986)[a]

Vendor	Furnace type[b]	Flue gas flowrates, acfm	Outlet PM emissions, gr/dscf at 12% CO_2[c]		
			0.03	0.02	0.01
C	MB	24,523	1,712	1,712	1,762
C	MB	245,230	5,262	5,262	5,624
G	MB	245,230	6,000	6,000	6,000

[a] Installed capital costs reported are the purchase costs for one unit multiplied by a 1.6 adjustment factor. Auxiliary equipment costs are not included.

[b] MB = mass-burn.

[c] Outlet grain loading from fabric filters.



August 1988 Dollars
Logarithmic Scale
y (min) = 10; y (max) = 100

Figure 2.4-1. Capital cost estimates of an SD/FF for a model MB facility, and RDF facility.[2]

Table 2.4-2 summarizes the capital cost procedures for single SD/FF units. These procedures are based on achieving a PM control level of 0.01 gr/dscf at 12 percent CO_2.[3] The equation was developed from Figure 2.4-1.

From Table 2.4-2, the total direct costs can be estimated for single units by knowing the inlet flue gas flowrate and the length of ductwork needed. The flue gas flowrate is based on 125 percent of the design flowrate to accommodate variations in feed waste composition and operating conditions.[4] To estimate the costs of multiple units, the direct costs of a single SD/FF unit including auxiliary equipment are multiplied by the number of units.

2.4.2.2 Indirect and Other Costs. To be consistent with established EPA methodology, the equations were adjusted to distinguish direct costs (i.e., purchased equipment and installation costs) from indirect capital costs (i.e., engineering costs, construction and field expenses, contractor fees, start-up and performance test costs). To separate these costs, indirect costs are assumed to be 33 percent of the direct capital costs.[5] Contingency is assumed to be similar to that applied to fossil-fuel fired boilers.[6] Interest during construction and working capital is not included for air pollution control devices.[7] Costs are reported in December 1987 dollars. The Chemical Engineering Plant Cost Index for all equipment was used to escalate costs from August 1986 dollars.

2.4.3 Operating Cost Procedures

Table 2.4-3 presents the procedure for estimating operating costs. In general, the references in this table have been used in previous EPA cost analyses.

The operating and maintenance labor requirements for SD/FF are based on those used in fossil fuel industrial boiler cost analyses and assume that operating and maintenance labor costs bases would be essentially the same for coal-fired industrial boilers and MWC facilities. However, the maintenance material cost for SD/FF systems applied to MWC facilities is usually lower than the cost for systems at coal-fired boiler facilities, since uncontrolled SO_2 emissions are much higher from coal-fired boilers. Because SO_2 concentrations are lower at MWC facilities, less concentrated slurries can be

TABLE 2.4-2. CAPITAL COST PROCEDURES FOR SD/FF FOR NEW MWC'S[a]

Total Direct Costs (December 1987 dollars)

Single SD/FF Unit[b]: Costs, 10^3 \$ = 8.053 $(Q)^{0.517}$

Ductwork[b]: Costs, 10^3 \$ = $[1.3868 * L * Q^{0.5}]/1{,}000$

Fan[b]: Costs, 10^3 \$ = $[1.8754 * Q^{0.96}]/1{,}000$

Multiple Units: Multiply the above costs by the number of units

Indirect Costs = 33% of total direct costs

Contingency = 20% of sum of direct and indirect costs

Total Capital Costs = Total Direct Costs + Indirect Costs + Contingency Costs

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[a] Q = 125 percent of the actual flue gas flowrate, acfm
L = Duct length, feet

[b] Assumes that the total installed costs are 133 percent of the direct capital costs.

TABLE 2.4-3. ANNUAL OPERATING COSTS PROCEDURES FOR SPRAY DRYER/FABRIC FILTER FOR NEW MWC's[a]

	Reference
Operating Labor: 4 manhours/shift; $12/manhour	8, 9
Supervision: 15% of operating labor costs	10
Maintenance:	
Labor: 2 manhours/shift; 10% wage rate premium over operating labor wage	9
Materials: 2% of direct capital costs	11
Bag Replacement:	
Bags: $1.35/$ft^2$ for teflon-coated fiberglass; 2-year life for SD/FF; Bag replacement cost not included for SD only	12
Electricity: Cost Rate = $0.046/kwh	
Fan: 12.5 inches of water pressure drop	13, 14
Atomizer: 6kW/1,000 lbs/hr of slurry feed + 15kW	15
Pump: 20 feet of pumping height 10 psi discharge pressure 10 ft/sec velocity in pipe	16
Compressed Air: 2 scfm air/1,000 acfm flue gas; $0.11/1,000 scfm of air	17
Water: Calculate water flowrate required for cooling the flue gas to 300°F; water cost = $0.50/1000 gal	18
Lime: Based on lime feed rate calculated for a given stoichiometric ratio; lime cost = $70/ton	19
Solid Waste: Calculate solid waste collected by the spray dryer and fabric filter using PES program and apply appropriate ash disposal fee in $/ton; Assume $25/ton	20

(continued)

TABLE 2.4-3. (Continued)

	Reference
Overhead: 60% of the sum of all labor costs (operating, supervisory, and maintenance) plus materials	21
Taxes, Insurance, and Administrative Charges: 4% of total capital costs	21
Capital Recovery: 15-year life and 10% interest rate	22

[a] All costs are in December 1987 dollars.

used to achieve the same removal efficiency, which in turn result in less erosion of equipment and potential for plugging. Therefore, the maintenance material cost was estimated at 2 percent of the direct capital cost.[23] Estimating the material cost at 2 percent of the direct capital cost corresponds to 1.25 percent of the total capital costs.

The costs of bag replacement assumes a 2-year bag life, which is a typical bag-life for FF's.[24] A gross air-to-cloth ratio of 3:1 is used. Electricity costs include electricity consumed by the I.D. fan, atomizer, and slurry pumps. Electricity consumed by the I.D. fan is calculated using a pressure drop of 12.5 inches of water across the SD/FF. Atomizer electrical requirements are based on the amount of slurry feed.[25] Slurry pumping requirements are estimated from assumed pumping height, discharge pressure, and fluid velocity in pipe used in previous cost analysis.[26] The costs for compressed air for pulse-jet FF's are estimated from the air usage rate of 2 scfm/1,000 acfm of flue gas.[27] The stoichiometric ratio (moles of calcium per mole of SO_2 and HCl in the inlet flue gas) assumed is 2.5 to achieve 90 percent SO_2 and 97 percent HCl removals.

All cost rates are based on December 1987 dollars. The operating labor wage rate used is the average from those in the Department of Commerce, Survey of Current Business for private nonagricultural payrolls and EPRI's Technical Assessment Guide.[28,29] Electricity rates are from the Energy Information Administration, Monthly Energy Review.[30] The freight-on-board (FOB) costs for quick lime (calcium oxide, CaO), $45/ton bulk, are from the Chemical Marketing Reporter;[31] an additional cost of $25/ton is assumed for transportation, based on a hauling rate of $0.05/ton-mile and a 500-mile hauling distance. For estimating ash disposal costs, a tipping fee of $25/ton is used. For new plants, the operating costs will be based on the assumption of 8,000 hours of operation per year; however, operating costs can be calculated for any number of operating hours.

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2.5 COMPLIANCE MONITORING

Continuous emission monitoring (CEM) systems are used to determine compliance with emission limits for MWC facilities. The following sections describe monitoring systems for opacity, SO_2, HCl, O_2, and CO_2. Section 3.1 discusses the types of combustion control monitors required for good combustion practices.

2.5.1 Overview of Technology

2.5.1.1 Continuous Opacity Monitoring[1]. Stack opacity can be continuously measured using emission measurement systems based on the principle of transmissometry. Transmissometry measures the attenuation of visible light by particulate matter in stack effluent. Light from a lamp source is projected across the stack to a light sensor. The degree of attenuation (opacity) reflects the amount of light adsorbed and scattered by the particulate matter in the effluent.

The EPA regulations (Appendix B of 40 CFR Part 60) require the opacity monitoring system to operate for a minimum of 168 hours within certain performance specifications without unscheduled maintenance, repair, or adjustment. The regulations set forth minimum performance criteria for the following system parameters: calibration error ($\leq$3 percent), 24 hour zero drift ($\leq$2 percent), 24 hour calibration drift ($\leq$2 percent), and response time (10 seconds maximum). During or before installation, it is necessary to calibrate, zero, and span using calibration filters and to perform all alignments.

To validate accuracy, as required in 60.13(d), instruments automatically perform simulated zero and span calibration checks at selectable intervals (usually daily). It is also usually necessary to have an air purge system to prevent accumulation of particulate from condensing on the optical surfaces. Maintenance is typically required on an as-needed basis (usually weekly). This involves cleaning all filters, checking the optical alignment and the air purge system, and recalibrating the instrument.

2.5.1.2 Continuous SO_2 Monitoring[2]. Continuous monitoring of SO_2 emissions is typically accomplished by irradiating a given volume of sample air by ultraviolet (UV) or infrared (IR) light and measuring either the energy absorbed or the resulting fluorescence of the SO_2 molecules. Commercially available units differ in design and method, but in general it is necessary to: (1) collimate the light from the original source to provide a narrow band; (2) prepare the sample for analysis; and (3) increase the signal-to-noise ratio of the final signal via phase-sensitive detection, second derivative spectroscopic measurement, or other techniques.

All SO_2 monitoring systems required under NSPS must have complete zero and span calibration checks performed daily. If not, a weekly manual check is recommended. About every month, it is necessary to clean, service, and readjust the instrument. The actual maintenance schedule needed depends on the instrument and the site of application. Instruments utilizing filters, chillers, sample dryers, or support gases typically require more maintenance.

Continuous monitoring systems must be installed at sampling locations where representative measurements can be made of the total emissions from the affected facility, or can be corrected so as to be representative. The SO_2 monitoring system must be capable of operating for a 168-hour minimum within certain performance specifications without unscheduled maintenance, repairs or adjustments. The regulations (Appendix B of 40 CFR Part 60) set forth minimum performance criteria for the following system parameters: accuracy (<20 percent) and 24-hour calibration drift (2.5 percent of span). The calibration drift is determined using calibration gases (i.e., gases of known concentrations), gas cells, or optical filters. The relative accuracy is determined by measuring pollution concentrations with EPA reference methods while concurrently operating the continuous monitoring system.

2.5.1.3 Continuous HCl Monitoring. The EPA has not published performance specifications for HCl monitors, but is currently evaluating the reliability, accuracy, and reproducibility of various monitoring systems. The outcome of this evaluation will determine which monitoring systems will serve as the basis for any ensuing EPA performance specifications for continuous HCl monitors.

In brief, four types of extractive monitors are being evaluated and are available commercially. The first type is a wet chemical batch process.[3] A sample of the flue gas passes through an automatic bubbler system, and the HCl-laden liquor is sprayed against a specific ion electrode. The second type is a nondispersive infrared (NDIR) analyzer.[4] This instrument determines the HCl concentration of the sample flue gas by ratioing the peak heights of the flue gas and reference gas. Both types of monitors are certified for CEM applications in West Germany.

The third type uses a tape sampler.[5] A sample of gas is exposed to a chemically impregnated tape. The HCl in the flue gas reacts with the chemical on the tape leaving the tape stained. The instrument determines the HCl concentration by measuring the reduction in transmissivity of the tape. The last type of monitoring system is based on continuous spectrophotometry.[6] A sample of flue gas is contacted with a thiocyanate reagent stream in a column. The reagent leaving the column, which contains adsorbed HCl, is fed to the spectrophotometer to obtain an HCl signal. These two types are not certified for CEM applications in West Germany.

2.5.1.4 Diluent (O_2/CO_2 Monitoring).[7] Diluent monitors are an integral part of an SO_2 or HCl continuous monitoring system. Diluent concentrations (O_2 or CO_2 on a percent basis) are required to convert actual concentrations of SO_2 or HCl to concentrations at either 7 percent O_2 or 12 percent CO_2.

Continuous monitoring of O_2 is based on the paramagnetic properties of O_2 molecules and their response to nonhomogeneous magnetic fields or by oxide cell differential voltages. Monitoring CO_2 is accomplished through infrared absorption methods.

2.5.2 Compliance Monitoring Costs

Table 2.5-1 summarizes the continuous monitoring costs associated with PM only, acid gas only, and PM and acid gas controls combined. Except for HCl and operating costs for SO_2 and O_2 monitors, the monitoring costs are the same as those used by EPA in developing NSPS for both small and industrial steam generation units.[8] Costs for HCl monitors and operating costs for a combined SO_2/O_2 monitor are based on recent information.[8,9]

The capital costs were updated to December 1987 dollars using the *Chemical Engineering* Plant Cost Index for all equipment, while the operating costs were updated to the same time bases using the Bureau of Labor Statistics' *Producer Price* Index for all industrial commodities. An automatic data reduction system is included in all options shown in Table 2.5-1.

TABLE 2.5-1. CONTINUOUS MONITORING COST SUMMARY[10,11]
(December 1987 Dollars)

Pollutant	Method	Capital Costs ($1,000)	Operating Costs ($1,000/yr)[b]	Annualized Costs ($1,000/yr)[c]
PM	Opacity[a]	61	8	16
Acid Gas	SO_2 (inlet and outlet)	67	10	19
	HCl (inlet and outlet)	140	74	92
	O_2/CO_2	19	15	18
	Data Reduction System	31	4	8
	Total	256	103	137
PM + Acid Gas	Opacity[a]	61	8	16
	SO_2 (inlet and outlet)	67	10	19
	HCl (inlet and outlet)	140	74	92
	O_2/CO_2	19	15	18
	Total	286	107	145

[a] Includes costs for automatic data reduction system.

[b] Based on 2 certifications/year and maintenance requirements of 0.5 man-hour/day for opacity and O_2/CO_2 monitors and 1 man-hour/day for SO_2 and HCl monitors.

[c] Annualized costs include annual operating costs and capital charges on equipment and installation costs. Capital charges are based on a 15-year equipment life at 10 percent interest rate.

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3.0 PROCEDURES FOR EXISTING PLANTS

This section presents procedures for estimating costs for existing municipal waste combustion (MWC) plants. Most procedures presented in this section rely on those procedures discussed for new plants. However, additional procedures are developed which are unique to existing plants such as costs for combustion modifications to the combustors, flue gas cooling using humidification, and downtime associated with either the installation of the air pollution control device (APCD) or modifications to the combustor. This section also provides a methodology to assess the higher costs of installing APCD's at existing plants, compared to new plants, using retrofit factors.

Section 3.1 presents procedures for estimating costs for operating the existing combustor. Procedures for estimating costs of combustion modifications are presented in Section 3.2. Section 3.3 provides the procedure for estimating costs for flue gas temperature control using humidification. Sections 3.4, 3.5, and 3.6 discuss estimation of costs for particulate matter control, dry sorbent injection, and spray drying, respectively. Section 3.7 present the methodology to determine retrofit factors and additional site-specific costs. Downtime costs associated with the installation of an APCD or modifications to the combustor at an existing plant are discussed in Section 3.8.

3.1 OPERATION OF THE EXISTING COMBUSTORS

No capital costs are estimated for the combustors and other equipment associated with the balance of plant, because these costs are sunk and are independent of the costs for retrofitting additional APCD's. Therefore, only the operating costs of the combustors and the balance of the plant are considered. Operating costs procedures for new combustors and the balance of plant are presented in Section 2.1 and are assumed to be the same for existing plants. For existing plants, capital recovery costs are not included in the total operating costs.

3.2 COMBUSTOR MODIFICATIONS

3.2.1 Introduction

This section describes the methodology and assumptions used to estimate capital and annual costs associated with combustor modifications needed to ensure good combustion for MWC's.[1,2] The organization of this chapter is as follows:

- Section 3.2.2 discusses the approach used to estimate capital costs for each of the combustion modifications, including all assumptions. An example calculation is provided for each retrofit.
- Section 3.2.3 provides a methodology for estimating annual costs for MWC plants.

The Chemical Engineering Plant Cost Indices are used to convert costs to December 1987 dollars.

3.2.2 Capital Cost Procedures

Capital cost estimates were calculated for each retrofit component and expressed as a direct, installed cost, unless otherwise noted. When uninstalled equipment costs are provided, an installation factor is applied:

Direct Capital Cost (DCC) = 1.45(Equipment Cost)

The installation factor applies to delivered equipment in a solids processing plant.[3]

Capital costs that may vary based on unit size must be scaled using factors. For example, the cost of a modification, C, at a unit of a given size is scaled for a unit of different size by the following equation:

$$C_1 = C_2\,(TPD_1/TPD_2)^n$$

where:

C_1 = scaled capital cost of equipment at unit #1;

C_2 = capital cost of equipment at unit #2;

TPD_1 = capacity (tons per day) of unit #1; and

TPD_2 = capacity (tons per day) of unit #2.

The exponent n varies according to the retrofit application. It is assumed that the volumetric heat release (Btu/ft^3-hr) is constant for similar combustor types (i.e., mass-burn waterwall, RDF-fired, etc.). Therefore, for a given design, unit firing capacity (tons per day) scales directly with furnace volume. Consequently, a change in a given design feature will vary as the cube root of each resulting change in dimension modifications, and the exponent is 0.667. In the case of retrofitting a row of overfire air nozzles, where a one-dimensional change is required (along the width of the combustor), the exponent is 0.333. Perry's Chemical Engineers' Handbook also applies typical exponents for various pieces of equipment.[4] The exponent values range from 0.30 to 1.00 depending on the specific equipment. As noted in Perry's Chemical Engineer's Handbook, use of exponents to estimate costs results in a slightly higher probable error (10 to 50 percent) than study estimates (up to 30 percent).

Indirect capital costs (ICC) and contingencies must be applied to the direct capital costs (DCC) estimates.[5] Indirect capital costs, which include general facilities and engineering and home office costs, etc., are calculated as 30 percent of DCC:

$$ICC = 0.30(DCC).$$

A single contingency is applied to the DCC:

$$\text{Contingency} = 0.20(DCC).$$

The 20 percent contingency factor is applied in all cases except when a retrofit is judged to be especially difficult, such as with stoker (grate) replacement; a contingency factor of 30 percent is used in this case. The total plant capital cost (TPC) is calculated as follows:

TPC = DCC + ICC + Contingency.

The following subsections describe the costing methodology for specific retrofit elements, including:

- Stoker rehabilitation,
- Refractory-wall furnace reconfiguration,
- Fuel feeding modifications,
- Underfire air modifications,
- Overfire air modifications,
- Monitoring/control modifications,
- Auxiliary fuel burner installation, and
- Economizer installations for flue gas temperature reduction.

3.2.2.1 Stoker Rehabilitation

This modification includes demolition and replacement of existing stoker, drives, siftings hopper, siftings conveyor, and structural steel. It is also assumed that a new stoker is equipped with a ram feeder.

- Chesner reports direct capital costs for stoker rehabilitation for four 250-tpd units to be $4,160,000 (in December 1984 dollars) based on quotes from two stoker equipment suppliers.[6]

- Assume single unit cost for 250-tpd unit is $1,040,000.

- Apply CEP index:

  12/84 - 324.3
  12/87 - 332.5
  Unit Cost = 1,040,000 (332.5/324.3) = $1,066,000.

- Apply scaling factor and account for number of units:

  DCC = 1,066.000 (TPD/250)$^{.677}$(number of units).

- Example: Estimate the direct capital cost of replacing traveling grates with new reciprocating grates in two 375-tpd units:

  $$DCC = 1{,}066{,}000\ (375/250)^{.677}\ (2) = \$2{,}797{,}000.$$

3.2.2.2 Refractory-Wall Furnace Reconfiguration

This modification includes material and labor for reconstructing the combustion chambers and refractory-lined flues, including structural steel and refractory brickwork. It is assumed that new overfire air nozzles and sampling ports are included in the new furnace design.

- Chesner reports direct capital costs for furnace reconfiguration for four 250-tpd units to be $6,072,000 (12/84 dollars).[6]
- Assume single unit cost for a 250-tpd unit is $1,518,000.
- Apply CEP index:

  12/84 - 324.3
  12/87 - 332.5
  Unit Cost = $1,518,000(332.5/324.3) = $1,556,000.

- Apply scaling factor:

  $$DCC = 1{,}556{,}000(TPD/250)^{.667}(\text{number of units}).$$

- Example: Estimate the direct capital cost of reconstructing two 120-tpd refractory wall combustors:

  $$DCC = 1{,}556{,}000(120/250)^{.667}(2) = \$1{,}903{,}000.$$

3.2.2.3 Fuel Feeding Modifications

Ram Feeder - This modification includes material and labor, including the hydraulic system, for a new ram feeder, plus any necessary modifications to the feed chute.

- Nashville Thermal reports 1979 direct capital (installed) costs of ram feeders (one dual ram for each of two 360-tpd units) to be $360,000.[7]
- Assume that the unit cost is $180,000 for dual rams and $90,000 for single ram. (Single rams can be used for grates with widths up to 8 feet.)

- Apply CEP Index:

  1979 (yearly average) - 247.6
  12/87 - 332.5
  DCC = 90,000(332.5/247.6) = $121,000 per ram feeder.

- _Example_: Estimate the direct capital cost of retrofitting one ram on each of two 120-tpd units with 8-foot wide grates:

  DCC = $121,000(2) = $242,000.

RDF Metered Feeder - This modification includes installing metered feeding modules, consisting of two hoppers, one ram feeder, and one variable-speed drive conveyor per module.

- Equipment cost = $150,000 per module.[8]

- Apply installation factor to obtain direct capital cost:

  DCC = $150,000(1.45) = $217,500 per module.

- _Example_: Estimate the direct cost of retrofitting metered feeding modules on two 300-tpd RDF-fired facilities. Assume two distributors per unit and one module per distributor:

  DCC = $217,500(2 modules/unit)(2 units) = $870,000.

3.2.2.4 Underfire Air Modifications

Segmented Underfire Air Supplies - This modification includes installing segmented, separately controllable underfire air plenums.

- Lavalin estimated the direct capital cost of five new underfire air plenums to be $153,000 Canadian (2/85) for the Quebec City Incinerator.[9]

  Assume cost for one plenum = $153,000/5 = $30,600.

- Convert to U.S. dollars:[10] $Canadian = 1.35 ($U.S.)

  $U.S. = 30,600/1.35 = $22,700 (2/85 dollars).

- Apply CEP Index:

  2/85 - 325.4
  12/87 - 332.5

(22,700)(332.5/325.4) = 23,200.

- Apply scaling factor:

  (Quebec City is a 250-tpd unit.)

  $DCC = 23,200(TPD/250)^{.667}(h)(\text{number of units})$,

  where h = number of plenums.

- Example: Estimate the direct capital cost of installing a single underfire air plenum to the drying grate section of two 120-tpd units:

  $DCC = 23,200(120/25)^{.667}(1)(2) = \$28,400.$

Underfire Air Preheat - This modification includes a natural gas burner sized to provide sufficient heat input to raise combustion air temperatures from 68°F to 300°F.

- Example: Determine the size and direct capital cost of an auxiliary fuel burner required to preheat underfire air supplied to the drying grate. Assume that the unit size is 250 tpd.

  250 tpd(2000 lb/ton)(day/24 hr)(hr/60 min) = 347 lb/min MSW.

  Assume that the combustor operates at 150 percent excess air and that stoichiometric air requirements are 3.25 lb air/lb waste.

  Total air requirements are:

  (347 lb/min)(3.25 lb air/lb waste)(2.5) = 2820 lb air/min.

  Assume that 70 percent of total air is supplied as undergrate air, and 20 percent of undergrate air is supplied to the drying grate.

  (2,820 lb/min)(.70)(.20) = 395 lb/min at 68°F.

  $Q = mc_p T$

  where: Q = heat input,
  m = 395 lb/min (mass flowrate),
  c_p = 0.24 Btu/lb°F (specific heat of air at standard conditions), and
  T = 300 - 68 = 232°F.

  $Q = (395 \text{ lb/min})(0.24 \text{ Btu/lb}^{\circ}\text{F})(232^{\circ}\text{F})(60 \text{ min/hr}) = 1.32\ 10^6 \text{ Btu/hr}$

  Use a 1.4×10^6 Btu/hr burner.

- MITRE reports capital costs of burners ranging from capacity of 9.2×10^6 to 1.5×10^6 Btu/hr to be $1200 per burner.[11]
- Apply CEP Index:

  1981 (yearly average) - 297.0
  12/87 - 332.5

  1,200 (332.5/297.0) = 1340.
- Apply installation factor to obtain direct capital cost:

  DCC = 1,340(1.45) = $1,950 per burner.

3.2.2.5 Overfire Air Modifications

Flow modeling/thermal analysis studies are required in most cases prior to modifying overfire air systems. Overfire air modifications made at refractory-wall MWC's and tube and tile waterwall MWC's will usually require only new ducting, dampers, and nozzles. New overfire air rows in membrane-wall MWC's are assumed to require installation of new waterwall tube panels.

Flow Modeling/Thermal Analysis Studies[12] - These analyses include flow visualization studies, mixing and dispersion measurements, and flow distribution studies on a built-to-scale physical model. In addition, mathematical modeling is included as part of the thermal analysis.

Cold flow modeling	- $75,000
Thermal analysis	- $50,000
Total	$125,000

Ducting and Dampers -

- Ducting Capital Costs:[13] $C = 1.1(L)(Q)^{0.5}$,

  where L = Length (ft) and

  Q = 125 percent of the actual flue gas flowrate (acfm).
- Example: Estimate direct capital costs of ductwork and dampers required to supply overfire air to two rows of nozzles. Assume a gas flowrate of 21,400 acfm. Assume that the overfire air system

is designed to provide 40 percent of total air flow (8,560 acfm). At standard conditions, Q = 1.25(8,560) = 10,700 acfm.

Assume ducting length requirements are 100 feet.

$C = 1.1(100)(10,700)^{.5}$ = \$11,400(equipment cost).

- Damper Capital Costs: Chemical Engineering, December 29, 1980 presents cost curves for rectangular dampers.[14]

- Estimate costs of a damper to install in ducting. Assume that the damper is manually controlled and has a 1.5 ft^2 cross-sectional area. The damper equipment cost is \$400 (in December 1977 dollars).

Apply CEP Index:

12/77 - 210.3
12/87 - 332.5

400(332.5/210.3) = \$600 per damper (equipment cost).

Total equipment cost = Ducting costs + damper costs

\$11,400 + 600 = \$12,000.

Apply installation factor:

Total DCC = 1.45(12,000) = \$17,400.

Insulation for Ducting - Capital costs for ducting insulation vary from 3.5 to 22 percent of direct capital costs for ducting. Selection of the appropriate factor is based on flue gas temperature.[15]

- Example: Assume that ducting carries preheated air at a temperature of 300°F and that capital costs for the ducting are \$20,000. Estimate direct capital costs of insulation.

Perry's Chemical Engineers Handbook (Table 25-51) specifies a range of 3.5 to 6 percent of ducting costs over \$17,000. Select 6 percent as conservative number.

C = 20,000(0.06) = \$1,200.

Apply installation factor:

DCC = 1.45(1,200) = \$1,740.

Membrane Wall Overfire Air Nozzle -

- Lavalin reports direct capital costs for one row of nozzles installed at Quebec City Incinerator to be $40,000 (Canadian 2/85 dollars).[16]

- Convert to U.S. dollars:

  $U.S. = $Canadian/1.35
  $U.S. - 40,000/1.35 = 29,600 (2/85 dollars).

- Apply CEP Index:

  2/85 - 325.4
  12/87 - 332.5

  DCC = 29,600(332.5/325.4) = $30,200 per row.

- Apply scaling factor:

  (Quebec City is a 250-tpd unit.)

  $DCC = 30{,}200(TPD/250)^{.333}$(number of rows)(number of units).

- Example: Estimate direct capital costs for two new overfire air rows per unit for two 1000-tpd combustors:

  $DCC = 30{,}200(1000/250)^{.333}$(2 rows/unit)(2 units) = $192,000.

3.2.2.6 Combustion Controls and Monitors

Fully Automatic Combustion Controller - This modification includes all hardware and software required for converting a manual combustion control system to a fully automatic control (programmable logic controller).

- Direct capital costs for one unit are $200,000.[17,18]

- Additional units can be installed in control scheme using the same hardware. Incremental capital costs are restricted to those costs required for installation. Assume that the direct capital cost of an automatic controller for more than one combustor is:

  DCC = 200,000[1 + 0.45(N - 1)],

  where N = number of combustors, and
  installation factor = 45 percent of equipment costs.

  DCC = 200,000 + 90,000(3 - 1) = $380,000.

Monitors - Display readouts and data loggers are included for each monitor. Air flow monitors are venturi flow meters with pressure transducers.

- Direct capital cost of in situ CO/O_2 monitors - \$45,000[19].
- Direct capital cost of in situ CO monitor - \$22,000[19].
- Direct capital cost of air flow pressure monitors for underfire air plenums and overfire air headers - \$1,500 per plenum or row of overfire air nozzles.[20]

Oxygen Trim Control - This modification includes installation of a control loop which adjusts underfire air flowrate and/or plenum distribution based on feedback signals from an O_2 analyzer.

- Hampton, VA plant manager reports direct capital costs to be \$25,000 for two 100-tpd units.[21]
- Assume that these costs are fixed, per unit costs:

  DCC = \$12,500/combustor.

3.2.2.7 Auxiliary Fuel Burner Installation

- Gas pipeline costs:

  DCC = \$50,000 per 1/2 mile[22].

- Auxiliary gas burners - Capital costs of dual-fuel burner packages, including blowers, igniters, safety panels, and controls, are available for the following burner sizes.[23] An installation factor of 45 percent is applied to obtain direct capital costs.

Burner size (Btu/hr)	Equipment Cost	Direct Capital Cost
10.5	\$16,000	\$23,200
30.0	\$25,500	\$37,000
45.0	\$35,000	\$50,800
60.0	\$42,000	\$60,900

Burner equipment costs for sizes other than those provided above should be extrapolated based on size, and the 45 percent installation factor should then be applied.

- Example: Estimate the capital cost of providing auxiliary fuel to a facility with three 300-tpd combustors. Assume the nearest source of gas is one mile away, and each combustor requires two burners, each rated at 35 x 10^6 Btu/hr.

DCC of pipeline = \$100,000 and
Cost of one 35 x 10^6 Btu/hr burner = \$31,400.
Apply installation factor:
DCC = 1.45(31,400) = \$45,500.

Total direct capital costs for burners = \$45,500 and
(2 burners/unit) (3 units) = \$273,000.
Total direct capital costs = 100,000 + 273,000 = \$373,000.

3.2.2.8 Carbon Monoxide Profiling

This activity includes two days labor for three men in the field plus travel and reporting. Sampling is assumed to include O_2, carbon monoxide (CO), and temperature measurements in a 16-point array under six variable air distribution settings. Carbon monoxide profiling is required on only one combustor when multiple units of identical design are in place:

DCC = \$10,000 (Reference 24).

3.2.2.9 Economizer for Flue Gas Temperature Control

This modification includes a separate economizer module designed to reduce flue gas temperatures from 600°F to 450°F, along with the addition of ducting and a bypass damper.

- Equipment cost = \$45,000 (1986 dollars) for an economizer sized to handle flue gases from four 75-tpd units (300-tpd total).[25]
- Apply CEP Index:

  1986 - 318.4
  12/87 - 332.5

  45,000(332.5/318.4) = \$47,000.
- Apply installation factor:

  DCC = \$47,100(1.45) = \$68,100.
- Apply scaling factor:

  DCC = $68,100(TPD/300)^{.59}$.

- Example: Estimate the direct capital cost of installing one economizer for three 50-tpd units:

$$\text{DCC} = 68{,}100(150/300)^{.59} = \$45{,}200.$$

3.2.3 Operating Cost Procedures

Total annual costs include annual operating and maintenance (O&M) costs and annualized capital costs. Table 3.2-1 presents a summary of inputs used to estimate annual costs. The costs provided for each plant are incremental O&M costs. For example, if a plant is equipped with auxiliary fuel burners at baseline, it is assumed that the fuel is used for start-up and shutdown, and no incremental O&M cost is applied to the plant for auxiliary fuel consumption. Plants without auxiliary burners in place will incur additional costs for fuel consumption. The following examples illustrate the calculation of annualized costs associated with combustion controls.

- Example: A mass-burn refractory-wall MWC consisting of three 250-tpd combustors must add auxiliary fuel burners and operate the burners during start-up and shutdown. The facility maintains a five per week operating schedule. Determine the size of burners required to provide 60 percent of rated thermal load and estimate natural gas consumption costs.

$$\text{Combustor thermal load} = \frac{(250 \text{ ton/day})(2000 \text{ lb/ton})(4500 \text{ Btu/lb})}{(24 \text{ hr/day})}$$

$$= 94 \times 10^6 \text{ Btu/hr}$$

Assume for a refractory-wall facility that gas is fired for six hours during start-up and two hours during shutdown. Assume that the plant operates 50 weeks/year, and start-up/shutdown occurs weekly.

$$\text{Total gas use} = (50 \text{ wk/yr})(6 + 2 \text{ hours})(56 \times 10^6 \text{ Btu/hr})(3 \text{ units})$$
$$= 67.2 \times 10^9 \text{ Btu/yr}$$

TABLE 3.2-1. O&M COST INPUTS (DECEMBER 1987 DOLLARS)

Item	Value
Direct Operating Costs	
Operating Labor	\$12.00/hour
Supervision	15 percent of operating labor
Maintenance Labor	110 percent of operating labor (assume 1 hr/shift for maintenance of controls and monitors)
Maintenance Materials	100 percent of maintenance labor
Natural Gas	\$4.50 per 10^6 Btu
Water	\$0.50 per 1000 gallons
Steam	\$5.30 per 1000 lb
Solid Waste Disposal	\$25 per ton
Indirect Operating Costs	
Overhead	60 percent of all labor costs (operating, supervisory, and maintenance) plus 60 percent maintenance materials
Taxes, Insurance, and Administrative Charges	4 percent of total plant capital costs
Capital Recovery	15 year life and 10 percent interest rate $\text{CRF} = \frac{i(1 + i)^n}{(1 + i)^n - 1}$ where i = interest rate and n = number of years $\text{CRF} = \frac{.1 + (1.1)^n}{(1.1)^{15} - 1} = .1315$

Using a gas cost of $\$4.50/10^6$ Btu:

$C = (67.2 \times 10^9 \text{ Btu/yr})(4.50/10^6 \text{ Btu}) = \$302,000/\text{yr}$

- Example: Determine the annual costs for a combustion retrofit at the plant in the above example. Total plant capital costs are assumed to be $500,000, including installation of CO and O_2 monitors.

Direct Costs:

Assume 1 hr/shift (3 hr/day) maintenance of monitors and controls.

Maintenance Materials = (3 hr/day)(5 day/wk)(50 wk/yr)($13.20/hr) = $10,000/yr,

Maintenance Materials = $10,000/yr (100% of maintenance labor)

Gas costs = $302,000/yr, and

No additional operating labor is required.

Total Direct Annual Costs = 10,000 + 10,000 + 302,000 = $322,000.

Indirect Costs:

Overhead = 0.6(maintenance labor + maintenance materials).

Overhead = 0.6(20,000) = $12,000.

Taxes, Insurance and Administrative Charges = .04(total plant capital costs) = .04(500,000) = $20,000.

Annualized capital = .1315(500,000) = $66,000 assuming 15 year facility life and 10 percent weighted cost of capital.

Total Indirect Annual Cost = Overhead + Taxes, Insurance, and Administrative + Annualized Capital

= $12,000 + $20,000 + $66,000 = $98,000.

Total annual cost = Direct Cost + Indirect Cost

= $322,000 + $98,000 = $420,000/yr.

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3.3 HUMIDIFICATION

3.3.1 Overview of Technology

Humidification is used to cool the flue gas entering the particulate matter (PM) control device. Humidification can be used separately or in combination with dry sorbent injection. The primary objective of cooling is to reduce the temperature of the flue gas entering the PM control device to below that at which post-combustion formation of dioxin is suspected to occur (approximately 450°F).

The quantity of water required is a function of the temperature, flowrate, and moisture content of the flue gas at the inlet to the humidification chamber and the temperature reduction required.[1]

$$Q_w = (T_i - T_o) * Q_s * (1 - WTR/100)/940 \qquad (1)$$

where: Q_w = water required for flue gas cooling, lb/hr;
T_i = inlet flue gas temperature, °F;
T_o = outlet flue gas temperature, °F;
Q_s = flue gas flowrate, scfm; and
WTR = moisture content of the inlet flue gas, volume percent.

Flue gas temperatures at the combustor exit for refractory-wall combustors generally ranged from 1,400 to 1,600°F; for waterwall combustors, temperatures ranged from 400 to 600°F.

For units already using quench towers for flue gas cooling (primarily refractory-wall systems without heat recovery), the water feed rate is increased to achieve the additional cooling. For units without an existing flue gas cooling system, a humidification chamber is installed. The humidification chamber diameter is sized for a flue gas velocity of 10 feet/second and a chamber length-to-diameter (L/D) ratio of 3 to 1.[2] To minimize PM fallout and impingement of wetted solids on chamber walls, no baffles or other internals are used. Pressure nozzles are used for water atomization.

A secondary effect of cooling the flue gas entering the PM control device is a reduction in flue gas volume (i.e., acfm) and a corresponding increase in the specific collection area (SCA) thereby enhancing the PM collection efficiency of the ESP. However, because MWC ESP's operate at temperatures above the temperature of maximum particle resistivity (300 to $400^{\circ}F$ for most fly ashes), decreasing flue gas temperature may in some instances increase fly ash resistivity enough to create ESP back corona problems and impair PM collection efficiency. Because of the current lack of information on resistivity-temperature relationships for MWC fly ash, this analysis assumes that humidification does not alter particulate resistivity enough to cause ESP operating problems. As a result, the impact of humidification on ESP performance is estimated based solely on the change in SCA due to flue gas volume reduction.

3.3.2 Capital Cost Procedures

Capital costs are estimated for existing facilities without an existing flue gas cooling system. Direct capital costs include the humidification (evaporative cooling) chamber including the vessel and supports, water spray system and controls, and duct modifications. Direct equipment cost for the humidification chamber are based on the flue gas flowrate using the following equation:[3]

$$\text{Equipment Costs (\$)} = 0.372 * Q + 67,980 \qquad (2)$$

where: Q is 125 percent of the actual inlet flue gas flowrate (acfm) to accommodate variations in waste composition and operating conditions.[4]

The costs estimated by equation 2 are in December 1987 dollars. Originally, the costs were in December 1977 dollars and were adjusted to December 1987 dollars using the Chemical Engineering Plant Cost Index for all equipment. The equipment costs are then adjusted for retrofit difficulty based on the procedures described in Section 3.7.1.

Costs for instrumentation, taxes, freight, and installation are estimated using indirect cost factors for venturi scrubbers.[5] The resultant procedure for estimating capital cost is summarized in Table 3.3-1.

3.3.3 Operating Cost Procedures

Table 3.3-2 presents procedures for estimating operating and maintenance (O&M) costs for the humidification chamber. Because of the simple design and operating requirements of the system, O&M labor and maintenance materials are assumed to be at the low end of those presented in Reference 6 (i.e., using the wet scrubber labor and materials requirements). Other O&M costs include water and the electricity used by the pumps. All costs are based on December 1987 dollars. An operating labor wage of the $12/hr was used. This wage was the average of the labor wages reported by both the Department of Commerce Survey of Current Business for private nonagricultural payrolls and EPRI's Technical Assessment Guide for utility power plants.[7,8] The labor wage reported by EPRI in January 1985 dollars was updated to December 1987 dollars using the Bureau of Labor Statistics' Producer Price Cost Index for all industrial commodities, prior to averaging. An electricity cost of $0.046/kWh was obtained from the Energy Information Administration Monthly Energy Review.[9] Equipment life is assumed to be 15 years.

TABLE 3.3-1 CAPITAL COST PROCEDURES FOR HUMIDIFICATION[10,11]

Equipment Costs (December 1987 dollars)

1. Humidification Chamber and Pumps:[a]

   Cost, \$ = $0.372 * Q + 67{,}980$

2. Ductwork

   Cost, \$ = $0.981 * L * Q^{0.5}$

Retrofit Purchase Equipment Costs = 1.18 * Equipment Costs * Retrofit Factor (from Section 3.7)

Installation Direct Costs = 0.56 * Purchased Cost

Indirect Costs[b] = 0.35 * Purchased Cost

Total Capital Costs = Purchased Equipment Costs + Installation Direct Costs + Indirect Costs
= 1.91 * Purchased Costs

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[a] Q = 125 percent of the actual flue gas flowrate, acfm
L = Duct length, feet.

[b] Includes a contingency of 3 percent of the purchased costs.

TABLE 3.3-2 OPERATING AND MAINTENANCE COSTS FOR HUMIDIFICATION

		References
Operating Labor:	0.5 man-hours/shift; wages of $12/hr	6, 8
Supervision:	15% of operating labor costs	12
Maintenance Labor:	0.5 man-hours/shift 10% wage premium over operating labor wages	12
Maintenance Materials:	1% of total capital investment	13
Water:[a]	0.00012 * Q_w * (hours of operation) * (water costs, $/1000 gal) cost of $0.50/1000 gal	14
Electricity:[a,b]	1.587×10^{-4} * Q_w * (hours of operation) * (electricity costs, $/kWh) cost of $0.046/kWh	15
Overhead:	60% of the sum of all labor costs (operating, supervisory, and maintenance) and maintenance materials	15
Taxes, Insurance, and Administrative Charges:	4% of the total capital costs	15
Capital Recovery:	15-year life and 10% interest rate	

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[a] Q_w = water injection rate, lb/hr, (from Equation 1 in Section 3.3.1).

[b] Assume 20 feet of pumping height, 100 psi discharge pressure, and 10 ft/sec velocity in pipe.

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3.4 PARTICULATE MATTER CONTROL RETROFIT

This section discusses three electrostatic precipitator (ESP) control alternatives for reducing PM emissions from existing MWC facilities. These alternatives are: installation of a new ESP (discussed in Section 3.4.1), increasing the plate area of an existing ESP (Section 3.4.2), and rebuilding an existing ESP to improve performance (Section 3.4.3).

3.4.1 Installation of a New ESP

The procedures for estimating ESP capital costs for new plants (described in Section 2.2.2) are applicable to the procedures used for existing plants. The existing plant cost procedures include site-specific retrofit factor and scope adders used to estimate the cost of demolition, replacement, relocation of existing equipment, new ducting, and stacks, if needed.

3.4.1.1 Capital Cost Procedures. The procedures developed for estimating the capital costs of ESP's for new plants (described in Section 2.2.4.1) are used to estimate the direct costs of major equipment, including the fans and ash handling. Estimated duct lengths are required to calculate duct costs for connecting the ESP to an existing plant. The estimated direct costs of new equipment and ducts are then multiplied by site-specific retrofit factors determined by the procedures described in Section 3.7.1.

Total direct capital costs for retrofit are calculated as the sum of the adjusted equipment costs plus any scope adders. Scope adders are additional significant costs for items, such as chimneys or demolition, that are required for an accurate estimate of the ESP retrofit. Determination of scope adder costs is described in Section 3.7.2.

After the total direct capital costs have been estimated, the remainder of the capital cost procedure (for indirect and contingencies costs) is the same as for ESP's installed in new plants as described in Section 2.2.4.1.

3.4.1.2 Operating Cost Procedures. Operating costs for retrofit ESP's are estimated using the same procedures as those for new plants discussed in Section 2.2.4.2. The costs of taxes, insurance, and administrative charges are estimated as a fraction of the total retrofit capital costs. The proposed procedures also allow operating hours to be varied to reflect model plant specifications.

3.4.2 Increase in ESP Plate Area

Additional ESP plate area is installed when the existing ESP is too small to achieve the desired PM control. Addition of plate area is accomplished by installing a new ESP in series with the existing ESP. This approach results in minimum facility downtime and will simplify cost estimation relative to the addition of plate area to the existing ESP.

3.4.2.1 Capital Cost Procedures. The procedures developed for estimating the capital costs of ESP's for new plants (described in Section 2.2.4.1) are used to estimate the direct costs of installing additional ESP plate area. First, the required particulate removal efficiency is calculated based on the PM emission limit desired and the inlet PM concentration. This removal efficiency is then used to calculate the required specific collection area (SCA) using either equation 2 or 4 presented in Section 2.2.4.1. Next, the SCA of the existing ESP is subtracted from the calculated SCA to determine the additional SCA required. The additional SCA required is used to calculate the additional plant area requirement and the direct costs of the second ESP using equations 1 or 3 in Section 2.2.4.1. The required duct length is estimated for each model plant based on the equipment configuration for that plant. The estimated direct costs of the new ESP and ducts are then multiplied by a site-specific retrofit factor determined according to the guidelines discussed in Section 3.7.1. Appropriate scope adders are costed based on procedures described in Section 3.7.2.

After the total direct capital costs have been estimated, the remainder of the capital cost procedure (for indirect and contingency costs) are the same as for ESP's installed in new plants presented in Section 2.2.4.1.

3.4.2.2 Operating Cost Procedures. Operating costs for the second ESP are estimated using procedures for new plants discussed in Section 2.2.4.2. Only those costs associated with the second ESP are included. Because operating, supervision, and maintenance labor are available for the existing ESP, it is assumed that no additional labor requirements are necessary to operate and maintain the second ESP.

3.4.3 ESP Rebuild

An ESP rebuild can be used with existing ESP's with PM removal efficiencies lower than those predicted in either Figures 2.2-2 or 2.2-4 for a new ESP with equivalent SCA. Rebuild of an ESP includes replacing worn or damaged internal components (plates, frame, and electrodes), upgrading controls and electrical systems for more effective energization, and flow modeling to evaluate gas distribution. The ESP rebuild does not include making design changes to the existing ESP, such as changes to the plate-electrode geometry or addition of collection area.

3.4.3.1 Capital Cost Procedures. The procedures developed for estimating the capital costs of ESP's for new plants (described in Section 2.3.4.1) are used to estimate the direct costs of ESP rebuild. Based on contacts with ESP vendors, a typical cost for rebuilding an existing ESP is roughly 30 percent of the total capital cost of a new ESP of equivalent size, but can be as high as 50 percent of the new ESP cost.[1,2]

The recommended procedure for estimating the total capital costs for ESP rebuild is to use 30 percent of the cost for a new ESP. This factor assumes equipment costs of 42 percent of the cost of a new ESP plus installation and indirect equipment cost multipliers of 0.33 and 0.27, respectively.[3] These indirect cost multipliers are lower than those used for new ESP's because: (1) new foundations, supports, piping, insulation, and painting are not required and (2) engineering and erection expenses are reduced relative to the costs for a new ESP. Site-specific retrofit factors are not used since the rebuild is performed within the existing ESP.

3.4.3.2 Operating Cost Procedures. The operating and maintenance costs after ESP rebuild are the same as before the rebuild with the exception of additional waste removal. The additional waste removal requirements are based on the incremental reduction of PM achieved after the ESP is rebuilt.

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3.5 DRY SORBENT INJECTION RETROFIT

3.5.1 Overview of Technology

Cost procedures are presented in Section 2.3 for the injection of dry sorbent into the furnace or duct of a new plant. The major distinctions between the design of sorbent injection systems for most existing facilities versus new facilities are (1) the reuse of an existing ESP rather than a new fabric filter for PM control, and (2) the higher capital costs to reflect the difficulty of a site-specific retrofit. For existing facilities not equipped with an ESP, new fabric filters can be used.

Another retrofit option for duct sorbent injection at existing facilities is to inject dry sorbent following an existing spray humidification chamber. In this option, the flue gas leaving the combustor is cooled by humidification to 350°F before it enters the ESP, or to 300°F in the case of a fabric filter. Dry sorbent is injected after the gas is humidified to minimize cake buildup in the duct.

3.5.2 Capital Cost Procedures

The procedures developed for estimating the capital costs of dry sorbent injection for new plants are used to estimate the direct capital cost of major equipment and ductwork for retrofit installations. Because the major equipment components of dry sorbent injection (reagent storage and handling system) can be located in remote areas, difficulties associated with spacial constraints (i.e., access/congestion) and underground obstructions is generally minimal. Based on the application of dry sorbent injection to coal-fired utility boilers, the direct capital cost for new plants is increased by 10 percent to account for the estimated costs of modifying an existing duct in the case of duct sorbent injection, modifying an existing overfire air system in the case of furnace sorbent injection, or modifying an existing humidification chamber.[1]

The total direct capital costs for retrofit also include the cost of any scope adders such as additional ducting or existing equipment demolition that is required to accurately estimate dry sorbent injection retrofit costs at a specific site. Additional ductwork can be estimated using cost equations in Section 2.3. Scope adders are defined in Section 3.7.2. Determination of scope adder costs is also described in Section 3.7.2.

After the total direct capital costs have been estimated, the remainder of the capital cost procedure for estimating indirect capital costs and contingencies is the same as for dry sorbent injection at a new plant presented in Section 2.3.4.1.

3.5.3 Operating Cost Procedures

Operating costs for retrofit dry sorbent injection installations are estimated using the same procedures discussed in Section 2.3.4.2 for new plants. Operating costs for existing plants are higher than for new plants of equivalent sizes because the maintenance expenses are affected by access and congestion difficulties. This increased cost is handled by calculating maintenance materials as a percentage of the total capital investment. The costs of taxes, insurance, and administrative charges are based on the total retrofit capital costs.

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3.6 SPRAY DRYER RETROFIT

3.6.1 Overview of Technology

Spray dryers (SD) combined with fabric filters (FF) can be retrofitted at existing plants where very high levels of CDD/CDF and acid gas control are required. Key technology considerations include reconfiguration of the ducting between the combustor outlet and stack, and the availability of space for installing sorbent handling equipment, SD vessel, FF, and ash disposal facilities.

Stand-alone SD costs were developed for this study to evaluate the costs of retrofitting a new SD in front of an existing particulate control device. Cost procedures presented in Sections 3.6.2 and 3.6.3 can be applied to estimate SD retrofit costs at existing plants. In most cases, the existing particulate control device is an ESP. For cases where it is determined that additional plate area is required to handle the increase in fly ash loading to the ESP caused by the SD, costing procedures presented in Sections 2.2 and 3.4.2 for modifying ESP's to add plate area should be used.

3.6.2 Capital Cost Procedures

The procedures developed for estimating the capital costs of SD/FF systems for new plants (described in Section 2.4.4) can be used to estimate the direct capital cost of major equipment and ducts for retrofit installations. Required duct lengths are used to estimate the duct costs for connecting the SD system to an existing plant. The estimated direct costs of new equipment and ducts are then multiplied by site-specific retrofit factors determined by the procedures in Section 3.7.1.

Capital costs for stand-alone SD systems are based on quotes obtained from three manufacturers.[1-3] These quotes, shown in Table 3.6-1, exclude the costs of any particulate control device. As discussed in Section 2.4, direct capital costs were correlated with flue gas flowrates. The direct capital cost equation in Table 3.6-2 for a single SD unit was developed from these quotes. The correlation coefficient (R^2) for this equation is 0.81. Figure 3.6-1 shows the relationship of both the predicted SD direct capital costs and the vendor costs with flue gas flowrate. The accuracy of the

TABLE 3.6-1. VENDOR QUOTES FOR SPRAY DRYER DIRECT CAPITAL COSTS (in 1000$ August 1988)

Vendor	Combustor Type[a]	Combustor Size, tpd[b]	Flue gas Flowrate, acfm	Direct Capital Costs
A	MB/WW	100	24,000	890
A	MB/RC	250	49,000	1,225
A	RDF	300	82,800	1,575
A	MB/WW	750	210,000	2,725
A	RDF	1,000	393,000	3,930
B	MB/WW	100	24,000	850
B	MB/RC	250	49,000	1,400
B	RDF	300	82,800	900
B	MB/WW	750	210,000	2,500
B	RDF	500	196,500	2,150
C	MB/WW	100	24,000	1,300
C	MB/RC	250	49,000	2,170
C	RDF	300	41,400	1,650
C	MB/WW	750	210,000	3,430
C	RDF	500	196,500	2,560

[a] MB/WW = mass burn/waterwall
MB/RC = mass burn/rotary combustor
RDF = refuse-derived fuel

[b] tpd = tons burned per day

TABLE 3.6-2. CAPITAL COST PROCEDURES FOR SPRAY DRYERS[a]

Total Direct Costs (December 1987 dollars)

Single SD Unit only: Costs, 10^3 \$ = $8.428 (Q)^{0.460} * N * RF$

Ductwork[b]: Costs, 10^3 \$ = $[1.3868 * L * Q^{0.5}]/1{,}000 * N * RF$

Fan[b]: Costs, 10^3 \$ = $[1.8754 * Q^{0.96}]/1{,}000 * N * RF$

Multiple Units: Multiply the above costs by the number of units

Indirect Costs = 33% of total direct costs

Contingency = 20% of sum of direct and indirect costs

Total Capital Costs = Total Direct Costs + Indirect Costs + Contingency Costs

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[a] Q = 125 percent of the actual flue gas flowrate, acfm
L = Duct length, feet
N = Number of units
RF = Retrofit factor, dimensionless

[b] Assumes that the total installed costs are 133 percent of the direct capital costs.

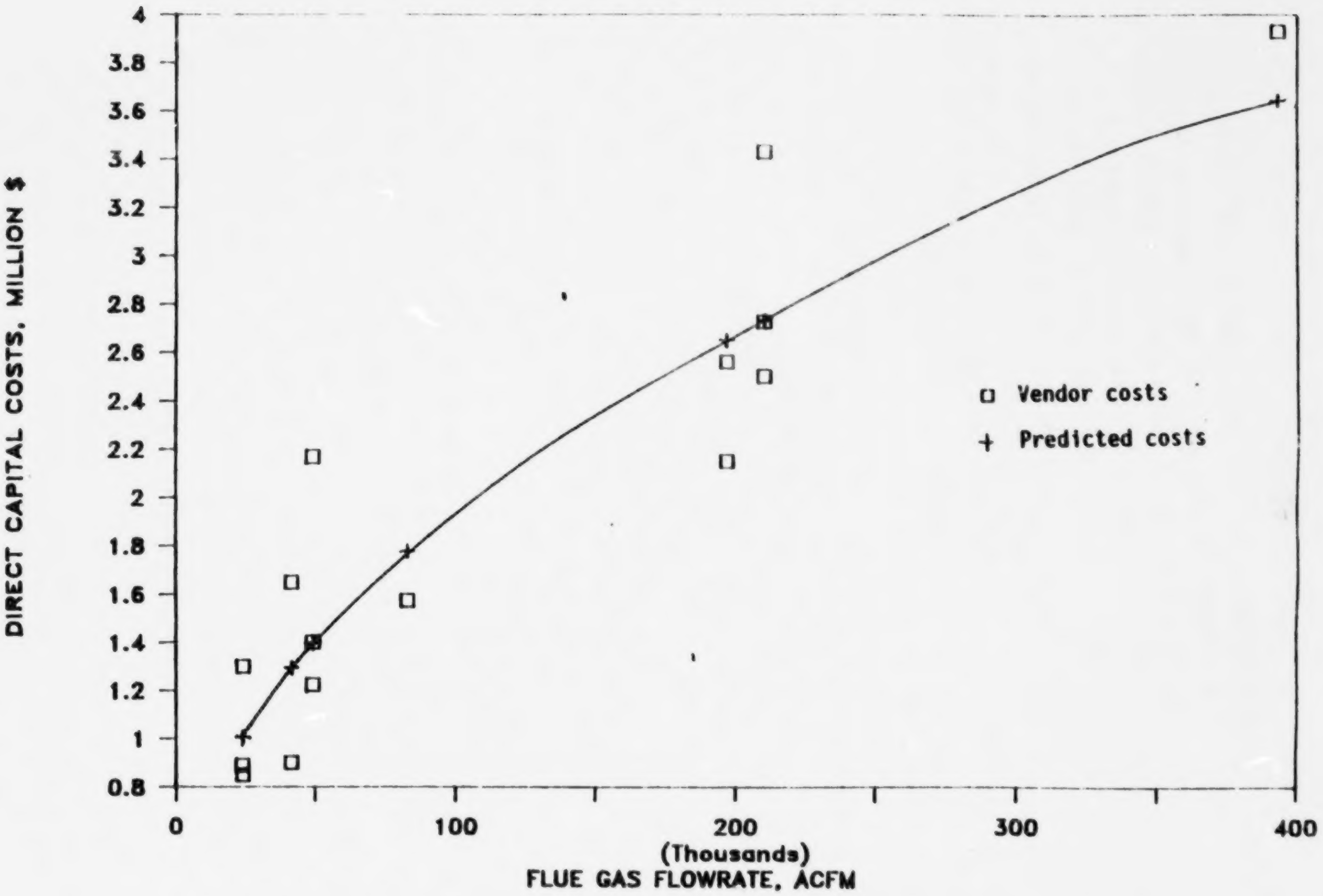

Figure 3.6-1. Correlation of SD direct capital costs (in August 1988 dollars) from the SD manufacturers and flue gas flowrate.

±30 percent. It should be noted that the costs shown in this figure are reported in August 1988 dollars and that the flue gas flowrate is the actual flue gas flowrate. The SD direct capital cost equation in Table 3.6-2 was derived by de-escalating the predicted cost curve shown in Figure 3.6-1 to December 1987 dollars using the Chemical Engineering Plant Index and by correcting for 125 percent of the actual flue gas flowrate. Comparing the direct capital costs for SD with those for SD/FF estimated using procedures in Section 2.4, the SD costs are generally between 50 and 60 percent of the costs for a SD/FF for flue gas flowrates ranging from 25,000 to 400,000 acfm. These flue gas flowrates cover the range of flowrates from small modular units to large RDF units. For ESP reuse, the costs of additional plate area, if any, estimated from procedures presented in Section 3.4.2 should be included.

The required duct length is estimated for each model plant based on the proposed air pollution control device (APCD) equipment configuration for that plant. The estimated direct costs of new equipment and ductwork are then multiplied by site-specific retrofit factors described in Section 3.7.1.

The total direct capital cost for retrofit is calculated as the sum of the adjusted new equipment costs plus any scope adders. Scope adders incorporate additional capital costs for items such as chimneys or demolition that are required for SD retrofit. Determination of scope adder costs is described in Section 3.7.2.

After the total direct capital cost has been estimated, the remainder of the capital costing procedure for indirect capital costs and contingencies is the same as for SD/FF installation at a new plant (see Section 2.4.2).

3.6.3 Operating Cost Procedures.

Operating costs for retrofit SD/FF installations are estimated using the same procedures as for new plants in Section 2.4.3. Table 3.6-3 presents the annual operating cost procedures for stand-alone SD's. Annual operating costs for the SD system alone exclude costs associated with the PM control device, such as bag replacement, compressed air, and solid waste costs. Operating labor, supervision, and maintenance labor costs for the SD alone are half those for a similar SD/FF system. Electricity costs for the I.D. fan are based on 5.5 inches of water pressure drop for an SD compared with

TABLE 3.6-3. ANNUAL OPERATING COSTS PROCEDURES FOR STAND-ALONE SPRAY DRYERS FOR NEW MWC's[a]

		Reference
Operating Labor:	2 man-hours/shift; $12/man-hour	4, 5
Supervision:	15% of operating labor costs	6
Maintenance:		
Labor:	1 man-hour/shift; 10% wage rate premium over operating labor wage	5
Materials:	2% of direct capital costs	7
Electricity:	Cost Rate = $0.046/kwh	
Fan:	5.5 inches of water pressure drop	4, 5
Atomizer:	6kW/1,000 lbs/hr of slurry feed + 15kW	8
Pump:	20 feet of pumping height 10 psi discharge pressure 10 ft/sec velocity in pipe	9
Water:	Calculate water flowrate required for cooling the flue gas to 300°F; water cost = $0.50/1000 gal	10
Lime:	Based on lime feed rate calculated by assuming a stoichiometric ratio of 1.5:1; lime cost = $70/ton	11
Overhead:	60% of the sum of all labor costs (operating, supervisory, and maintenance) plus materials	12
Taxes, Insurance, and Administrative Charges:	4% of total capital costs	12
Capital Recovery:	15-year life and 10% interest rate	13

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[a] All costs are in December 1987 dollars.

12.5 inches of water pressure drop for a SD/FF. Operating costs for ESP reuse are estimated from procedures presented in Section 3.4.2 for additional ESP plate area.

Operating costs for existing plants are higher than for new plants of equivalent size, since maintenance expenses will be affected by access and congestion difficulties. This increased cost is handled by calculating maintenance materials as a percentage of the total capital investment. The costs of taxes, insurance, and administrative charges are based on total retrofit capital costs. These procedures also allow operating hours to be varied to meet model plant specifications.

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3.7 DETERMINATION OF RETROFIT FACTORS AND SCOPE ADDER COSTS

The costs of air pollution control device (APCD) installation at an existing plant are greater than at a new facility due to higher construction costs imposed by site access and congestion, longer duct runs caused by space limitations, and the need to demolish and relocate some existing facilities. Procedures for estimating these costs at MWC's were adapted from procedures developed for the Electric Power Research Institute (EPRI) for retrofitting APCD's at existing electric generating plants.[1] These additional costs are divided into two types of adjustments: retrofit multipliers (discussed in Section 3.7.1) and scope adders (discussed in Section 3.7.2).

3.7.1 Retrofit Factors

Site-specific retrofit factors can be estimated based on access and congestion problems associated with retrofitting APCD's at existing plants. Depending on the level of accessibility and congestion, one of four factors (ranging from 1.02 to 1.42) is recommended based on the guidelines shown in Table 3.7-1. The total direct costs of new APCD equipment excluding ductwork are multiplied by this retrofit factor to estimate retrofit costs.[2]

3.7.2 Scope Adders

Scope adders are site-specific costs for additional ducting, chimneys, demolition, or any other major items that can be included in retrofit cost estimates in addition to the main control system equipment. Estimating procedures for some common scope adders are described here.

3.7.2.1 Ducting. Direct capital costs for ducts are estimated using the equation described in Section 2.2 for new plants. The duct costs are then multiplied by the retrofit factor from Section 3.7.1 to estimate the direct capital cost of ducts for existing plants. Depending on chimney and APCD tie-in difficulties at the model plant, the ductwork retrofit factor may be different than that chosen for the APCD.

3.7.2.2 Stacks. The installed capital cost of stacks is estimated from equations developed for industrial boilers.[3] Total direct and indirect capital cost data from one manufacturer were correlated into separate equations for lined and unlined stacks, and for stacks larger and smaller than

TABLE 3.7-1. SITE ACCESS AND CONGESTION FACTORS FOR RETROFITTING APCD EQUIPMENT AT EXISTING PLANTS[a]

Retrofit factor	Congestion level	Guidelines for selecting retrofit factor
1.02	Base Case	Interferences similar to a new plant with adequate crew work space. Free access for cranes. Area around combustor and stack adequate for standard layout of equipment.
1.08	Low	Some aboveground interferences and work space limitations. Access for cranes limited to two sides. Equipment cannot be laid out in standard design. Some equipment must be elevated or located remotely.
1.25	Medium	Limited space. Interference with existing structures or equipment which cannot be relocated. Special designs are necessary. Crane access limited to one side. Majority of equipment elevated or remotely located.
1.42	High	Severely limited space and access. Crowded working conditions. Access for cranes blocked from all sides.

[a] Reference 4.

5 feet in diameter (stacks larger than 5 feet in diameter and 100 feet tall are normally tapered). For a lined acid-resistant stack, the equations for direct and indirect capital cost, updated to December 1987 dollars, are:

$$\text{Cost, } 10^3\ \$ = [26.2 + 0.089 \times (H) \times (1 + 4.14\ D)] \text{ for } D \geq 5 \text{ ft and}$$
$$\text{Cost, } 10^3\ \$ = [26.2 + 0.080 \times (H) \times (1 + 4.33\ D)] \text{ for } D < 5 \text{ ft}$$

For an unlined stack, the equations are:

$$\text{Cost, } 10^3\ \$ = [26.2 + 0.0625 \times (H) \times (1 + 2.59\ D)] \text{ for } D \geq 5 \text{ ft and}$$
$$\text{Cost, } 10^3\ \$ = [26.2 + 0.087 \times (H) \times (1 + 2.20\ D)] \text{ for } D < 5 \text{ ft,}$$

where

H = stack height, ft and
D = stack diameter, ft.

To estimate the total capital costs, the direct and indirect costs are increased by 20 percent to account for contingency.

3.7.2.3 Demolition and Replacement. Costs for demolition of existing buildings required for installation of new APCD equipment are estimated according to EPRI guidelines.[5] In general, demolition cost is estimated by multiplying the amount of material to be demolished or moved (i.e., square feet of building space) by an appropriate cost factor in Reference 5. These estimates are made on a plant-specific basis as needed. Costs for demolition or replacement of existing equipment such as ductwork, fans, and ESP's are assumed to be the same as the costs for installing the same equipment.

REFERENCES

1. Stearns Catalytic Corporation. Retrofit FGD Cost-Estimating Guidelines. Prepared for Electric Power Research Institute. Palo Alto, CA. Publication No. CS-3696. October 1984.

2. Reference 1. pp. 4-1 to 4-3.

3. Bowen, M.L. and M.S. Jennings (Radian Corporation). Costs of Sulfur Dioxide, Particulate Matter, and Nitrogen Oxides Controls on Fossil Fuel-Fired Industrial Boilers. Prepared For the U. S. Environmental Protection Agency. Research Triangle Park, NC. Publication No. EPA-450/3-82-021. August 1982. p. 2-11.

4. Reference 1. p. 5-4.

5. Reference 1. pp. 4-9 to 4-14.

3.8 DOWNTIME COSTS FOR RETROFIT MODIFICATIONS

In many situations, the retrofit equipment cannot be installed during a normally scheduled maintenance shutdown and thus will result in additional downtime and loss of MWC revenues during retrofit. The loss of revenue is mainly from: (1) a loss of steam and/or electrical sales and (2) a loss of tipping fees from receiving MSW. It is assumed that the work force at the facility would be productive during the downtime period and that the cost of idle workers can be ignored.

To estimate the downtime costs due to loss of revenue, the length of downtime required to install the APCD must be estimated. Table 3.8-1 presents ranges of unit downtimes required to apply combustion control and install various APCD's on existing MWC facilities. Once the downtime period is estimated, Sections 3.8.1 and 3.8.2 present the procedures used to estimate costs for the loss of steam and electrical sales and the loss of tipping fees, respectively. Costs attributed to the loss of revenue are treated as a one-time cost that is annualized over the useful life of the APCD.

3.8.1 Procedures to Estimate Loss of Steam and Electricity Sales

3.8.1.1 Loss of Steam Sales. To estimate the costs of loss of steam during downtime, the amount of steam that would have been generated during the downtime period is multiplied by a sales price for steam (typically in dollars per 1,000 lb of steam). A typical steam price in December 1987 dollars is \$5.50/1,000 lb of steam.[1] For example, the lost revenues from steam sales for a facility normally producing 10,000 lb/hr of steam are \$1,320 per day (i.e., \$5.50/1,000 lbs steam times 10,000 lbs steam/hr times 24 hours).

3.8.1.2 Loss of Electricity Sales. The cost of lost electricity sales is estimated by multiplying the amount of lost electricity generation by the electricity price. The electricity price is assumed to be the same as the electrical cost rate used in this report to estimate APCD electricity costs (\$0.046/kWh in December 1987 dollars). Applying this procedure, the cost of lost electricity sales for a facility with a 1,000 kW capacity turbine is \$1,100 per day (i.e., \$0.046/kWh times 1,000 kW times 24 hours).

TABLE 3.8-1. DOWNTIME REQUIREMENTS IN MONTHS[a]

	Combustor downtime (months)
Combustion Modifications	0.25-4
ESP-Rebuild	1-2
ESP-Add plate area	0.5-1[b]
Retrofit Spray Dryer	1[b]
Retrofit Sorbent Injection	0.5-1[b]
Humidification	0.25-1

[a] Reference 2.

[b] If there are significant space limitations, up to an additional 6 months could be required.

3.8.2 Procedures to Estimate Costs from Loss of Tipping Fees

Downtime costs associated with loss of tipping fees are estimated by multiplying an appropriate tipping fee (typically $25/ton) by the increase in tonnage of solid waste disposal. The increase in solid waste is the amount of feed that would have been reduced in the combustor plus the fly ash that would have been collected by the existing PM control device, if the combustor were operating during the downtime period. For example, if the weight of MSW fed to a 100 tpd combustor is reduced by 75 weight percent during combustion (including bottom ash and fly ash), the tonnage of solid waste to be disposed would increase from 25 tpd during combustor operation up to 100 tpd when the unit is shut down. The increase in solid waste disposal costs is approximately $1,880, based on a $25/ton tipping fee (i.e., $25/ton times 75 tons per day).

REFERENCES

1. Electric Power Research Institute. TAG™-Technical Assessment Guide (Volume 1: Electricity Supply-1986). Palo Alto, CA. EPRI No. P-4463-SR. December 1986. p. B-4.

2. Memorandum from White, D.M. and J.T. Waddell, Radian Corporation, to R.E. Myers, EPA/ISB. June 3, 1988. Time Requirements for Retrofit of Particulate Matter (PM), Acid Gas, and Temperature Control Technologies on Existing Municipal Waste Combustors (MWC's).

APPENDIX B

DETAILED COST EQUATIONS